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ELECTRODIALYSIS IN ADVANCED WASTE TREATMENT

J. D. Smith, et al

Federal Water Pollution Control Administration
Cincinnati, Ohio

February 1967

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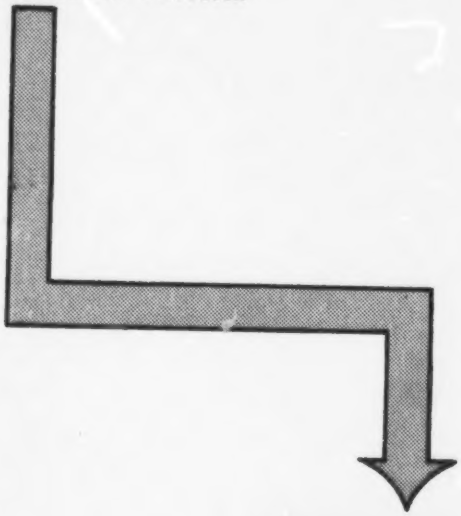

ELECTRODIALYSIS IN ADVANCED WASTE TREATMENT

U. S. DEPARTMENT OF THE INTERIOR
Federal Water Pollution Control Administration
Publication WP-20-AWTR-18

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ELECTRODIALYSIS
IN
ADVANCED WASTE TREATMENT

by

J. Douglas Smith and John L. Eisenmann

for

The Advanced Waste Treatment Research Activities
Research and Development
Cincinnati Water Research Laboratory
(Robert A. Taft Sanitary Engineering Center)

This report is submitted in fulfillment of Contract No. SAph 76690 between the Public Health Service and Ionics, Inc. Submitted February 1, 1965.

U.S. DEPARTMENT OF THE INTERIOR
Federal Water Pollution Control Administration
Cincinnati, Ohio
February 1967

In its assigned function as the Nation's principal natural resource agency, the United States Department of the Interior bears a special obligation to ensure that our expendable resources are conserved, that renewable resources are managed to produce optimum yields, and that all resources contribute their full measure to the progress, prosperity, and security of America - now and in the future.

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AWTR--Waste-Treatment Processes

Triplicate tear-out abstract cards are placed inside the back cover to facilitate information retrieval. Space is provided on the card for the user's accession number and for additional key words.

Water Pollution Control Research Series Publication No. WP-20-AWTR-18

WPCR SUB-SERIES

Waste-Treatment Processes (AWTR). A variety of methods have been used to remove part of the pollutants from wastewater before it is discharged. In many situations, existing methods are not adequate to provide the degree of treatment required to control pollution effectively. Research in waste treatment is being directed at developing methods ranging from lower-cost or higher-efficiency versions of conventional treatment technology to total wastewater purification systems capable of removing virtually all pollutants and even of producing reusable effluents. The Federal Water Pollution Control Administration's research in this area covers the entire field of waste treatment including treatment of industrial wastes, with special emphasis given to waste-treatment methods capable of total pollution control and purification of wastewater for deliberate reuse. This subseries, Waste-Treatment Processes, contains reports on results of this research by the Administration.

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AWTR-3	Ultimate Disposal of Advanced-Treatment Waste Part 1. Wet Oxidation Part 2. Incineration	999-WP-3
AWTR-4	Waste-Water Renovation Part 1. A Design Study of Freezing and Gas Hydrate Formation Part 2. Feasibility Tests of Freezing	999-WP-4
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AWTR-8	Ultimate Disposal of Advanced-Treatment Waste Part 1. Injection Part 2. Placement in Underground Cavities Part 3. Spreading	999-WP-10
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AWTR-11	Evaluation of the Use of Activated Carbon and Chemical Regenerants in Treatment of Waste Water	999-WP-13
AWTR-12	Evaluation of Various Adsorbents and Coagulants for Waste-Water Renovation	999-WP-14
AWTR-13	Electrochemical Treatment of Municipal Waste Water	999-WP-19
AWTR-14	Summary Report: The Advanced Waste Treatment Research Program. January 1962 through June 1964	999-WP-24
AWTR-15	Feasibility of Granular, Activated Carbon Adsorption for Waste-Water Renovation. 2.	999-WP-28
AWTR-16	Adsorption of Biochemically Resistant Materials from Solution. 2.	999-WP-33
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ABSTRACT

A bench-scale experimental study was made to determine the practicality of partially demineralizing municipal wastewater by electrodialysis. Using filtration alone and filtration followed by carbon adsorption as pretreatments, long-term runs were made with a municipal secondary effluent to determine scaling and fouling effects of the water upon the electrodialysis membranes. During these runs a study was made to determine how high a ratio of product rate to concentrate rate could be obtained without forming scale on the membranes. Some fouling of anion membranes occurred with filtration and carbon adsorption pretreatment; without carbon adsorption, fouling was much worse. Scale formation was not a problem at product to concentrate ratios less than ten. At times much higher ratios were obtained without difficulty. The results of these runs indicated that the total operating cost for a 10-mgd electrodialysis treatment plant might be less than 10¢ per 1,000 gal.

Information is presented on the relative selectivity of the membranes for the common ions in wastewater. Included are data on the long-term runs in which actual wastewater was used, and some data on laboratory prepared solutions. A theoretical discussion of selectivity is included.

SECTION I
INTRODUCTION

This report describes the second year of research conducted under Contract No. SAph 76690 for the United States Public Health Service Advanced Waste Treatment Research (AWTR) Program. The first year's work was broadly summarized in a paper[1] presented at the United States Public Health Service Seminar on Advanced Waste Treatment at Cincinnati, Ohio, on May 15, 1962.

During the first year of research, it was learned that alkylbenzene sulfonate could be removed from solution by ion-exchange and by electrodialysis by use of ion-exchange membranes. ABS did not readily pass through these membranes, however, but had to be eluted periodically with an acetone-HCl mixture. The porosity of the ion-exchange membranes had a pronounced effect upon ABS uptake, i.e., a low porosity or "tight" membrane did not take up ABS to the extent shown by a high porosity or "loose" membrane. This same phenomenon has been reported by others[2] and is probably due to the increased steric effect of the "tight" membranes.

Also during this first year, because asbestos fiber appeared to possess a definite adsorptive capacity for ABS, its use for ABS adsorption, both as a thin protective sheet adjacent to the anion permeable membranes in an electrodialysis stack and as a filter media in columns and centrifuges, was investigated.

By the beginning of the second year's work, other processes had been found that removed organic refractories with relative efficiency and economy[3,4,5]. Electrodialysis did, however, appear promising for the economical removal of inorganics; the second year of research, therefore, was concerned with this aspect of waste-water renovation. Several questions needed answers before electrodialysis costs in a waste-water renovation application could be accurately estimated. As in any industrial process, the unit product cost of an electrodialysis installation is composed of both capital investment and plant operating expenses. Since most of the capital investment consists of the initial and replacement membrane cost, some idea of the effective life of the membranes was necessary. Any performance decrement from normal brackish-water operations that might be inherent in a waste-water renovation application had to be determined so that this decrement could be anticipated in an accurate estimate of electrodialysis operating costs. These factors were investigated in the second year of research. The program consisted, primarily, of four phases:

1. Continuous electrodialysis demineralization of a municipal secondary effluent to study the effects of this effluent upon ion-exchange membrane life and general system performance.

2. A similar operation under more varied conditions of flow rates, current densities, product-to-waste ratios, etc., to obtain necessary engineering design information and to study the effects of increased waste-stream concentrations on system performance.

3. Membrane and system ionic selectivity studies designed to estimate relative rates of removal of the various ionic species present in treated municipal waste waters.

4. A preliminary design of a 10-mgd electrodialysis plant, and a cost estimate based upon the present state of development of commercially available electrodialysis equipment.

SECTION II
EXHAUSTION RUNS

It was suspected initially that fairly complete removal of suspended solids and dissolved ionizable organics would be required for the most efficient electrodialysis system performance. Experience in demineralizing brackish well waters had shown that suspended organic material tends to adhere to the surface of the membranes, forming a slime. Previous work[2] had shown that ABS would eventually foul anion-permeable membranes, which would result in an increase in electrical resistance and a decrease in membrane capacity. Evidence also exists that an operating electrodialysis stack may act as a culture medium for bacteria should any be present in the water[6].

Two series of exhaustion runs were made to evaluate the significance of these factors in a waste-water renovation application. Much of the experience from the first exhaustion series, described below, was reported at the American Chemical Society Symposium on Water Renovation in Cincinnati, Ohio, in January, 1963.[7]

A. EXHAUSTION - SERIES 1

1. Apparatus

The test apparatus illustrated schematically in Figure 1 was used in both series. System 1 was used to estimate membrane life and electrodialysis performance characteristics while demineralizing filtered municipal secondary effluent. A Fullflow SSD-10-3-3/8 filter* preceded the electrodialysis stack to remove suspended matter that might plug flow paths and adversely affect system performance. System II was used in a manner similar to System I. An activated carbon column was included in this system, however, to remove organic refractories (as measured by COD and ABS analyses) prior to the electrodialysis step. This column consisted of a section of 3-1/2-inch-diameter lucite tubing 24 inches long, capped at each end and filled with Pittsburgh Type OL-20x50-mesh activated carbon. Carbon was replaced when analysis indicated ABS breakthrough. Systems I and II were operated in parallel from the same secondary effluent source.

Secondary effluent was obtained on a semi-weekly basis from the Clinton Sewage Treatment Plant in Clinton, Massachusetts. This plant is a trickling-filter plant of approximately 4-mgd capacity and is about 5 years old. When visited on October 15, 1962, the Clinton plant was clean and well-maintained. Effluent was obtained twice a week, generally at 7 PM on Tuesdays and 11 AM on Saturdays to coincide with the time cycle of most highly concentrated effluent, and was delivered to a storage tank adjacent to the research laboratory.

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*Mention of commercial products does not imply endorsement by the Federal Water Pollution Control Administration.

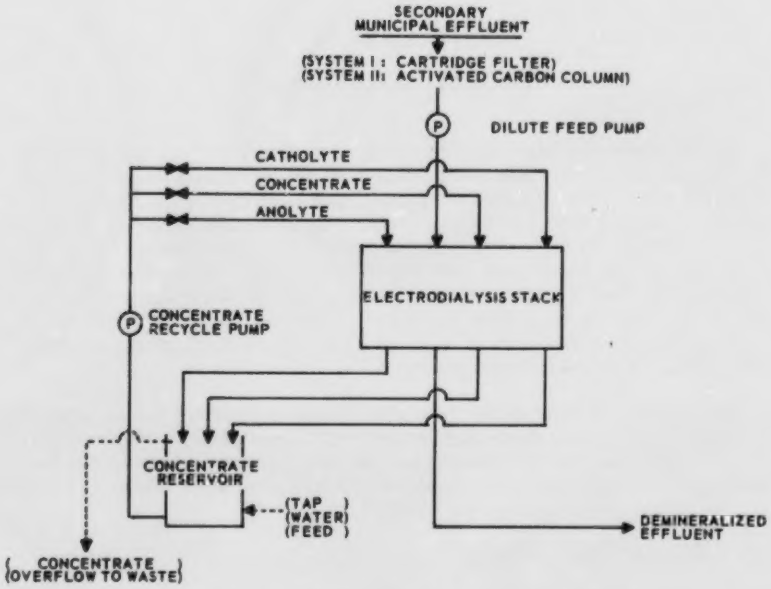

SECONDARY
MUNICIPAL EFFLUENT
(SYSTEM I: CARTRIDGE FILTER)
(SYSTEM II: ACTIVATED CARBON COLUMN)
P
DILUTE FEED PUMP
CATHOLYTE
CONCENTRATE
ANOLYTE
ELECTRODIALYSIS STACK
P
CONCENTRATE
RECYCLE PUMP
CONCENTRATE
RESERVOIR
(TAP
(WATER)
(FEED)
DEMINERALIZED
EFFLUENT
(CONCENTRATE
(OVERFLOW TO WASTE)

Figure 1. Flow diagram for exhaustion series 1 and 2 (dashed line indicates modifications for series 2).

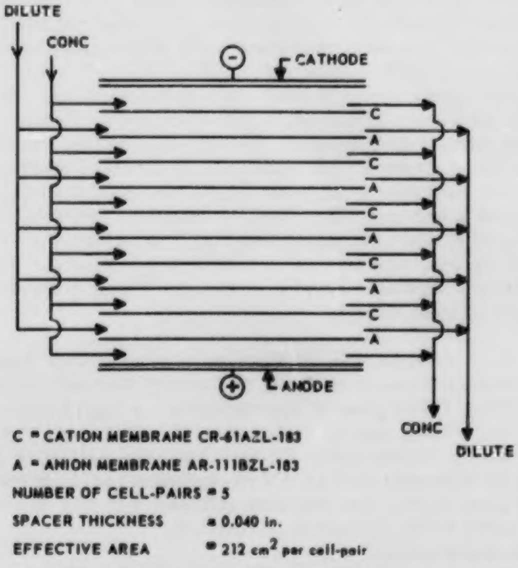

DILUTE
CONC
−
CATHODE
C
A
C
A
C
A
C
A
C
A
+
ANODE
CONC
DILUTE
C = CATION MEMBRANE CR-61AZL-183
A = ANION MEMBRANE AR-111BZL-183
NUMBER OF CELL-PAIRS = 5
SPACER THICKNESS = 0.040 in.
EFFECTIVE AREA = 212 cm² per cell-pair

Figure 2. Electrodialysis stack configuration.

Table 1 gives a typical analysis of Clinton effluent.

2. Preliminary Test - System I

Clinton effluent was pumped continuously for 2 weeks through a 10-cell-pair electrodialysis stack that used 9- by 10-inch ionics' Nepton ion-exchange membranes[8,9] (System I). At the end of that time, the stack was disassembled and the membranes visually examined. A brown scum or film was on the membranes, which probably was due to suspended matter in the sewage since there was no filter in the system during this trial. The membranes were washed in tap water, and the stack was rebuilt using the same membranes. The rebuilt stack consisted of only five cell-pairs. A stack configuration is shown in Figure 2.

Table 1. A TYPICAL ANALYSIS OF CLINTON, MASSACHUSETTS, SECONDARY SEWAGE EFFLUENT
January 24, 1963

Constituent	Concentration, mg/l	
Sodium, Na^+	34	
Potassium, K^+	12	
Ammonium, NH_4^+	22	
Calcium, Ca^{++}	12	
Magnesium, Mg^{++}	1	
Total cations		81
Chloride, Cl^-	29	
Bicarbonate, HCO_3^-	118	
Nitrate, NO_3^-	1	
Sulfate, $SO_4^=$	37	
Phosphate (ortho), $PO_4^{\equiv}$	14	
Phosphate (poly), $PO_4^{\equiv}$	14	
Total anions		213
Total inorganics		294
ABS	2.7	
COD	143	
Alkalinity, meq/l	1.94	
pH	6.9	
Temperature, °C	11	
Resistivity, ohm-cm	2680	

Exhaustion Runs

3. Exhaustion Test Results - Systems I and II in Parallel

Operating information on the first exhaustion series is tabulated chronologically in Appendix A. Concentrating streams were recirculated from 4-liter reservoirs in both Systems I and II. These reservoirs were drained and refilled with tap water daily. The cartridge filter was installed in the feed line ahead of the electrodialysis stacks.

a. System I

The electrodialysis stack in System I began operation on December 11, 1962, operating at constant voltage (5V). After 290 hours of continuous operation, the stack resistance had approximately doubled (100 to 193 ohms). At this time 2 liters of 0.5N HCl were pumped through the diluting compartments of the stack in an attempt to elute the materials fouling the membranes. This elutant was recovered and analyzed (Table 2); very large concentrations of iron, calcium, and magnesium were found.

Table 2. ANALYSIS OF 0.5N HCl ELUTION
SYSTEM I - DILUTE STREAM

December 23, 1962

Constituent	Concentration, mg/l
Sodium, Na^+	44
Potassium, K^+	28
Ammonium, NH_4^+	45
Calcium, Ca^{++}	248
Magnesium, Mg^{++}	61
Iron, Fe^{++}	133
Chloride, Cl^-	17,100
Bicarbonate, HCO_3^-	0
Nitrate, NO_3^-	9
Sulfate, $SO_4^=$	106
Phosphate (ortho + poly), $PO_4^{\equiv}$	31
ABS	2.85
Org. N	24

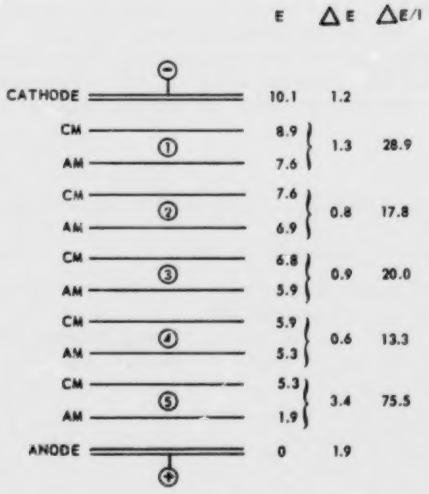
After the acid elution, the stack resistance returned to its original value (approximately 100 ohms). Following an additional 29 hours of operation, however, the resistance had increased to 185 ohms. The stack voltage was increased at this time to 10 volts and varied as necessary to maintain approximately a 50 milliamp current. The stack was operated in this manner continuously for 96 hours, during which time the resistance varied somewhat with feed concentration, but did not increase noticeably. Inorganic removals obtained at the end of this period are tabulated in Table 3.

Table 3. SYSTEM I - INORGANIC REMOVAL

December 30, 1962

Constituent	Concentrate stream mg/l	Influent, mg/l	Effluent, mg/l	%Removal
Sodium, Na^+	340	16	6	62
Potassium, K^+	84	5	2	60
Ammonium, NH_4^+		4	2	50
Calcium, Ca^{++}	214	10	4	60
Magnesium, Mg^{++}	60	2	1	50
Iron, Fe^{++}	0	0	0	
Chloride, Cl^-		21	12	43
Bicarbonate, HCO_3^-				
Nitrate, NO_3^-	2	1.5	1	33
Sulfate, $SO_4^=$	200	32	16	50
Phosphate (ortho + poly), $PO_4^\equiv$	14	4.4	4.0	9
ABS	0.3	0.22	0.24	--
COD		25	21	--
pH	8.3	7.5	7.1	
Resistivity, ohm-cm		3680	8100	
Temperature, °C		11	11	
Conductance at 25°C, Micromhos/cm		348	158	
Estimated Normality, meq/l		3.5	1.6	54

At this time also, probes were inserted between membranes in the electrodialysis stack, and the voltage drops across each cell-pair were measured (Figures 3 and 4). One-third of the total voltage drop was found to be isolated in a single cell-pair (number 5). It was suspected that the large resistance in this cell-pair was due to calcium salt pre-

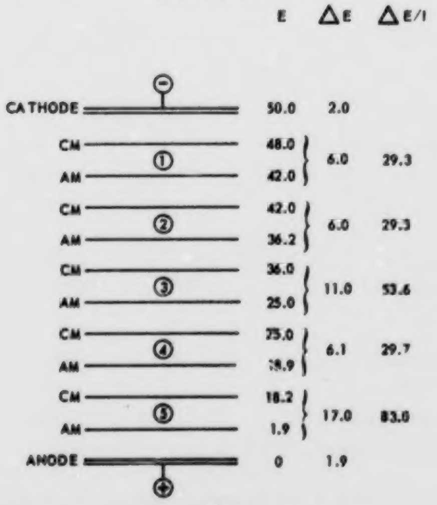

E
ΔE
ΔE/I
CATHODE
10.1
1.2
CM
8.9
①
1.3
28.9
AM
7.6
CM
7.6
②
0.8
17.8
AM
6.9
CM
6.8
③
0.9
20.0
AM
5.9
CM
5.9
④
0.6
13.3
AM
5.3
CM
5.3
⑤
3.4
75.5
AM
1.9
ANODE
0
1.9

I = 45 ma

DILUTE R_{OUT} = 5,950 ohm-cm

FLOW RATE = 120 ml/min

Figure 3. Voltage probes, system I – December 30, 1962 (E_T = 10V - before elution).


E
ΔE
ΔE/I
CATHODE
50.0
2.0
CM
48.0
①
6.0
29.3
AM
42.0
CM
42.0
②
6.0
29.3
AM
36.2
CM
36.0
③
11.0
53.6
AM
25.0
CM
25.0
④
6.1
29.7
AM
18.9
CM
18.2
⑤
17.0
83.0
AM
1.9
ANODE
0
1.9

I = 205 ma

DILUTE R_{OUT} = 11,200 ohm-cm

FLOW RATE = 120 ml/min

Figure 4. Voltage probes, system I – December 30, 1962 (E_T = 50V - before elution).

cipitation and flow path blockage within the concentrating compartment. Accordingly, 1.2 liters of approximately 1N HCl followed by 1 liter of tap water were pumped through the concentrating and electrode compartments. This elutant was recovered and analyzed (Table 4). Very large iron, calcium and magnesium concentrations were found, indicating both calcium-magnesium precipitation and iron fouling.

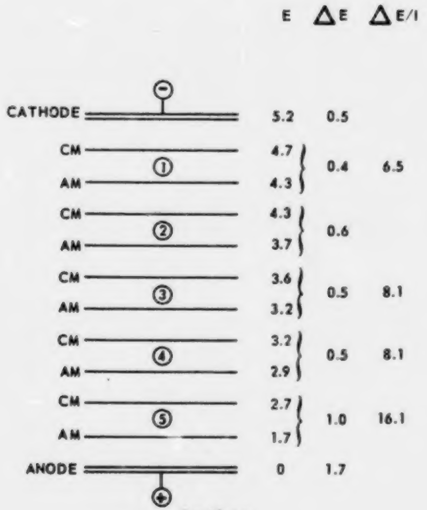
Table 4. ELUTANT ANALYSIS
SYSTEM I - CONCENTRATE STREAM

December 30, 1962

Constituent	Concentration, mg/l
Sodium, Na^+	92
Potassium, K^+	47
Ammonium, NH_4^+	0
Calcium, Ca^{++}	1,700
Magnesium, Mg^{++}	144
Iron, Fe^{++}	252
Chloride, Cl^-	19,650
Bicarbonate, HCO_3^-	0
Nitrate, NO_3^-	15
Sulfate, $SO_4^=$	185
Phosphate (ortho + poly), $PO_4^\equiv$	12
Org. N	105
ABS	3.6

After the concentrate-stream elution, the stack was again probed for cell-pair voltage drops (Figures 5 and 6). While some improvement was noted, a larger than average voltage drop was still noted in the number 5 cell-pair.

Stack voltage was increased to 25V and maintained at approximately this level for the next 55 hours. Current increased steadily during this period from an initial 98 ma to a final 205 ma; stack resistance decreased from 255 ohms to 118 ohms. A voltage probe at this time (Figure 7) showed a trend toward redistribution of cell-pair voltage drops indicating that calcium scale in concentrating compartments was being dissolved.

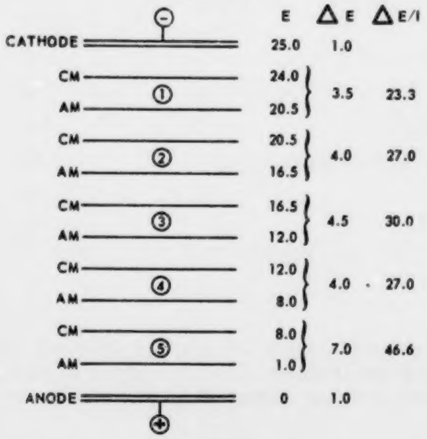

E
ΔE
ΔE/I
CATHODE
5.2
0.5
CM
①
AM
4.7
4.3
0.4
6.5
CM
②
AM
4.3
3.7
0.6
CM
③
AM
3.6
3.2
0.5
8.1
CM
④
AM
3.2
2.9
0.5
8.1
CM
⑤
AM
2.7
1.7
1.0
16.1
ANODE
0
1.7

I = 62 ma

DILUTE R_{OUT} = 9,270 ohm-cm

FLOW RATE = 120 ml/min

Figure 5. Voltage probes, system I – December 30, 1962 (E_T = 5V after elution).


E
ΔE
ΔE/I
CATHODE
25.0
1.0
CM
①
AM
24.0
20.5
3.5
23.3
CM
②
AM
20.5
16.5
4.0
27.0
CM
③
AM
16.5
12.0
4.5
30.0
CM
④
AM
12.0
8.0
4.0
27.0
CM
⑤
AM
8.0
1.0
7.0
46.6
ANODE
0
1.0

I = 150 ma

DILUTE R_{IN} = 6,170 ohm-cm

DILUTE R_{OUT} = 19,120 ohm-cm

T = 10.5°C

FLOW RATE = 120 ml/min

Figure 6. Voltage probes, system I – December 30, 1962 (E_T = 25V - after elution).

During the final 25 hours of operation, stack voltage was varied to maintain a current of approximately 120 ma. Typical inorganic removals during this period are tabulated in Table 5. After 502 total hours of operation, the electrodialysis stack was disassembled and membranes were examined.

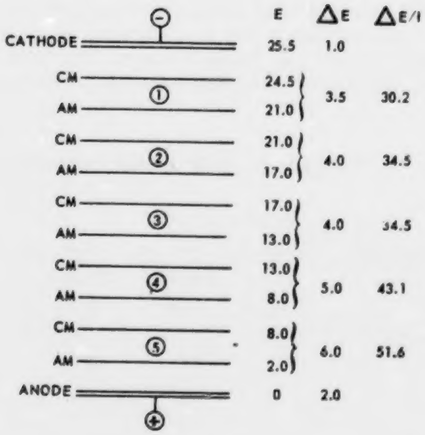
Table 5. SYSTEM I - INORGANIC REMOVAL

January 4, 1963

Constituent	Influent, mg/l	Effluent, mg/l	% Removal
Sodium, Na^+	50	20	60
Potassium, K^+	8	4	50
Ammonium, NH_4^+			
Calcium, Ca^{++}	10	18	-80
Magnesium, Mg^{++}	6	6	0
Iron, Fe^{++}			
Chloride, Cl^-	45	30	33
Bicarbonate, HCO_3^-	117	87	26
Nitrate, NO_3^-	6	1	87
Sulfate, $SO_4^=$	24	24	
Phosphate (ortho + poly), $PO_4^\equiv$	21	15	29
ABS	-	-	-
COD	257	107	58
Alkalinity, meq/l	1.92	1.42	26
pH	6.4	6.4	
Resistivity, ohm-cm	1690	3300	

Heavy slime deposits were found in all diluting compartments. $CaCO_3$ was found in one concentrating compartment. Iron was deposited on both electrodes.

Samples of the membranes were compared with samples of similar, but unused, membranes. The comparisons are shown in Table 6. The anion membrane sample showed some sign of capacity deterioration, probably due to sorption of ABS and other organic materials. Both membranes suffered an increase in resistivity probably from sorption of organic materials.


E
ΔE
ΔE/I
CATHODE
25.5
1.0
CM
①
24.5
21.0
3.5
30.2
AM
CM
②
21.0
17.0
4.0
34.5
AM
CM
③
17.0
13.0
4.0
34.5
AM
CM
④
13.0
8.0
5.0
43.1
AM
CM
⑤
8.0
2.0
6.0
51.6
AM
ANODE
0
2.0

I = 116 ma

DILUTE R_{OUT} = 14,400 ohm-cm

Figure 7. Voltage probes, system I – January 2, 1963 ($E_T = 25V$).

Table 6. SYSTEM I - MEMBRANE ANALYSIS

January 9, 1963

Property	Anion membrane AR-111BZL-183 Used	Anion membrane AR-111BZL-183 Unused	Cation membrane CR-61AZL-183 Used	Cation membrane CR-61AZL-183 Unused
Thickness, cm	0.065	0.073	0.066	0.068
Resistivity, ohm-cm^2	28.0	19.0	22.0	12.8
Water content, %	46.6	49.1	55.7	55.3
Capacity, meq/gm	1.29	1.78	2.78	2.75

b. System II

The electrodialysis stack in System II began operation December 13, 1962. It was operated parallel to the stack in System I, thereby serving more or less as a control to accentuate any performance decrement in

System I. This stack was probed for cell-pair voltage drops from time to time (Figure 8), and some calcium precipitation within at least one concentrating compartment appeared to take place. During the final 24 hours of operation, the System II stack resistance was consistently one-third that of System I. Table 7 shows an inorganic removal analysis during this final period.

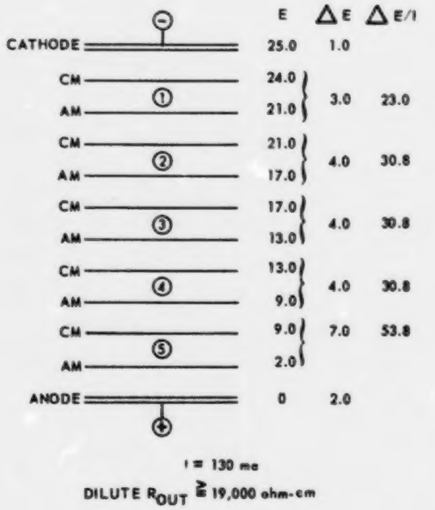

E
ΔE
ΔE/I
CATHODE
25.0
1.0
CM
24.0
①
AM
21.0
3.0
23.0
CM
21.0
②
AM
17.0
4.0
30.8
CM
17.0
③
AM
13.0
4.0
30.8
CM
13.0
④
AM
9.0
4.0
30.8
CM
9.0
⑤
AM
2.0
7.0
53.8
ANODE
0
2.0
I = 130 ma
DILUTE $R_{OUT} \geqq$ 19,000 ohm-cm

Figure 8. Voltage probes, system II – January 2, 1963 ($E_T = 25V$).

After 357 hours of operation, the electrodialysis stack was disassembled and examined. Very slight quantities of slime were found in the diluting compartments, most of which was adjacent to the anion membranes. Iron was deposited on both electrodes, although not nearly as much as that found in the System I stack. Heavy calcium deposits were found in two concentrating compartments. Samples of membranes were compared with similar, but unused, membranes (Table 8).

No decrease in capacity was noted in System II membranes; however, the increase in resistivities was similar to that for System I membranes.

Table 7. SYSTEM II - INORGANIC REMOVAL

January 4, 1963

Constituent	Influent, mg/l	Effluent, mg/l	% Removal
Sodium, Na^+	42	19	55
Potassium, K^+	8	4	50
Ammonium, NH_4^+			
Calcium, Ca^{++}	12	8	33
Magnesium, Mg^{++}	1	1	0
Iron, Fe^{++}	3	2	33
Chloride, Cl^-	44	30	32
Bicarbonate, HCO_3^-	128	54	58
Nitrate, NO_3^-	4	1	75
Sulfate, $SO_4^=$	26	24	8
Phosphate (ortho + poly), P		10	
ABS			
COD		55	
Alkalinity, meq/l	2.12	0.89	
pH	6.8	6.4	
Resistivity, ohm-cm	1850	4300	

Table 8. SYSTEM II - MEMBRANE ANALYSIS

January 9, 1963

	Anion membrane AR-111BZL-183		Cation membrane CR-61AZL-183	
Property	Used	Unused	Used	Unused
Thickness, cm	0.060	0.073	0.063	0.068
Resistivity, ohm-cm^2	25.0	19.0	25.0	12.8
Water content, %	49.2	49.1	55.7	55.3
Capacity, meq/gm	1.74	1.78	2.68	2.75

4. Discussion of Results

Performance parameters (stack resistance, removal efficiencies, etc.) were dependent upon feed concentrations. Comparison of the two systems, operating on essentially similar inorganic concentrations showed System I stack resistance to be consistently higher than that of System II during the latter stages of operation. System II did not noticeably deteriorate throughout 357 hours of total operation; stack resistance during the final 3 days of operation was, in fact, lower than the initial resistance. The increase in resistivity noted in both membrane analyses in both series could have been due in part to different equilibration conditions. The unused membranes were completely in the sodium and chloride form; membranes used during operation were equilibrated with the waste water being processed. Equilibration with 0.01N NaC1, prior to resistance measurements, resulted in a $C1^-$ or Na^+ concentration gradient from the surface to the center of the membrane sample. The centers of the samples may well have been in some form other than Na^+ or $C1^-$, which caused the apparent resistance increase. The greatest contribution to stack resistance was calcium deposits in the concentrating compartments. Table 5 shows that the calcium concentration leaving the stack in System I was higher than that entering, due to back-diffusion of calcium from a blocked concentrating compartment. Fouling of the cation membranes from excessive iron also occurred. This iron came from the unlined steel tank used for storage of the Clinton secondary effluent.

B. EXHAUSTION - SERIES 2

A second exhaustion series was undertaken to verify results from the first series, and to observe electrodialysis stack performance over a more prolonged period of operation.

1. Apparatus

Several modifications were made in the experimental apparatus before initiating this second series. The 275-gallon unlined steel storage tank was replaced by a Polyzinc/Rustoleum coated tank; this eliminated the iron fouling experienced in the first exhaustion series. The concentrating-stream plumbing was modified in both systems (as shown by the dashed lines in Figure 1) to allow the recirculating concentrate stream to be diluted continuously with tap water in a feed and bleed arrangement. This arrangement was expected to eliminate the calcium deposits experienced in the first exhaustion series without employing the customary acid treatment to the concentrate stream. Both electrodialysis stacks were rebuilt with new (but the same type: AR-111-BZL-183; CR-61-AZL-183) Ionics' Nepton membranes.

2. Test Procedure

Initially we hoped that secondary sewage effluent might be pumped continuously through the two parallel systems until their respective performances deteriorated significantly. It was soon apparent, however, that the volume of effluent obtained from Clinton was not sufficient to permit such operation without either using unrealistically low flow rates or increasing the number of effluent deliveries to three per week. Consequently, exhaustion series 2 became a succession of runs averaging approximately 50 hours each. Between runs, the electrodialysis stacks remained full of the previous run's feed. In this manner, the membranes in the stacks, although not continuously demineralizing, were continuously in contact with secondary effluent.

Product stream-flow rates through both systems were maintained at approximately 125 milliliters per minute. This corresponded to an in-stack flow-path velocity of 6.5 centimeters per second. Although current densities in both systems were varied at times to obtain data under extreme operating conditions (e.g., polarization), the current densities used throughout most of this series were about three-fourths the estimated limiting, or polarizing, current densities.

The cellulose, cartridge-type filters used in System I were changed weekly; the carbon adsorption column used in System II was changed initially when ABS breakthrough appeared. After the ABS concentration in the column effluent was monitored for several weeks, it became evident that a once-weekly carbon replacement would be sufficient to avoid ABS breakthrough. This procedure was followed, therefore, throughout the remainder of the System II exhaustion runs.

Electrodialysis stack operating parameters -- voltage, current, flowrate, pH of influent and effluent solution, temperature, and resistivity were monitored closely throughout the exhaustion series. These data and general operating information are tabulated chronologically in the form of an "operating log" in Appendix A. In addition, probes were inserted daily into the electrodialysis stacks in both systems, and voltage drops from membrane to membrane were recorded. The results of these voltage probes are in Appendix C.

To obtain demineralization information, grab samples from both the influent and effluent streams of both systems were taken twice weekly. These samples were analyzed for Na^+, K^+, NH_4^+, Ca^{+2}, Mg^{+2}, Cl^-, HCO_3^-, NO_3^-, $SO_4^=$, $PO_4^{\equiv}$, ABS, COD, alkalinity, organic nitrogen, pH, resistivity, temperature, and turbidity. The results of these analyses are tabulated in Appendix B.

Analyses of the Clinton secondary effluent showed that it was consistently deficient in nitrates, phosphates and ABS. Consequently, after April 3, 1963, this effluent was augmented with approximately 3 ppm ABS, 20 ppm NO_3^- and 30 ppm $PO_4^{\equiv}$.

Whenever the various parameters being monitored indicated some obvious decrement in the performance of a system, that system was shut down, the electrodialysis stack was disassembled, and the condition of the membranes was noted. The most common difficulties involved the build-up of slime or calcium scale, or both, on the anion permeable membranes. When necessary, the slime was washed off with tap water, and the scale was dissolved with HCl. The system was then reassembled and placed in operation.

During the final 350 hours of exhaustion series 2 the operation was modified to obtain preliminary information on the effects of increased waste-to-product concentration ratios. The feed and bleed dilutions of the concentrate reservoirs were discontinued while the concentrating streams continued to be recycled from the 4-liter reservoirs. These streams were permitted to increase in concentration for 24 hours. At the end of each 24-hour period, the reservoirs were emptied and refilled with 4 liters of tap water. Three to five drops of concentrated H_2SO_4 was added to this tap water, which resulted in an initial concentrate-stream pH of between 3.5 and 5.5. During the 24-hour recycle period, an additional 2 to 5 milliliters of concentrated H_2SO_4 was added to the concentrate reservoirs. Final concentrate-stream pH's at the end of the recycle periods were approximately 7.0. The product stream during the final 350 hours was about 180 liters per day; thus this type of operation involved essentially a 45 to 1 product-to-waste-volume ratio at the end of each 24-hour period. Actual operating conditions were, however, more severe than would normally be experienced with a 45 to 1 continuous-flow ratio. The waste stream pH's were higher during the latter stages of the 24-hour recycle periods than would be obtained with continuous flow systems. Consequently, some scaling problems were experienced, primarily in the cathode compartments.

At the end of exhaustion series 2 (approximately 1500 hours total operation over a 4-1/2-month period) both stacks were disassembled and examined. Calcium scaling was found in the cathode compartment of System I. Small amounts of slime were found in both diluting and concentrating compartments. Scaling was found in the cathode compartment and adjacent concentrating compartments of System II, the scaling increasing in severity with proximity to the cathode. Slime deposits found in System II diluting compartments were negligible.

3. Test Results and Discussion

The exhaustion runs were designed to evaluate generally the effects of a secondary effluent upon the performance of an electrodialysis demineralizing system, particularly the effects of this effluent upon the properties of the semi-permeable ion-exchange membranes in such a system.

From the information gathered during these exhaustion runs, data were extracted on the following parameters: (a) electrical resistance, (b) current efficiency, (c) energy consumption, (d) limiting current density, (e) ionic selectivity, and (f) membrane properties. Not all of these parameters are independent of each other; however, together they result in a fairly comprehensive picture of electrodialysis performance in the demineralization of a secondary sewage effluent.

a. Electrical Resistance

Many adverse effects of operation with secondary effluent are reflected as an increase in the electrical resistance of the electrodialysis stack. Any fouling of the membrane surfaces tends to increase this resistance. Scaling, i.e., calcium deposits, in the concentrating compartments increases the electrical resistance of the stack. Slime, or organic deposits, in the diluting compartments tends to, in effect, increase the thickness of the boundary films adjacent to the membranes. Since these films are more depleted in concentration than the bulk diluting solution, an increase in film thickness results in an increase in electrical resistance across the electrodialysis stack.

The electrical resistances of the five cell-pairs (anode and cathode voltages not included in computations) of both systems are tabulated in Table 9. Although these resistances vary daily because of day to day concentration changes, a general comparison between systems may be made on any specific date because of the similarity in the two feed-solution concentrations.

Table 9. ELECTRICAL RESISTANCE
EXHAUSTION RUNS, Series 2

	5-cell resistance (ohms)		Dilute solution R_{ave} (ohm-cm)		
Date, 1963	System I	System II	System I	System II	Remarks
2/11	137.9	172.3	5940	4460	Both systems apparently operating polarized.
2/18	121.5	146.7	4240	4180	
2/20	185.7	182.7	3720	3355	
2/26	124.1	372.7	3400	3220	
2/27	111.5	377.3	3275	3325	
3/7	39.5	36.4	4110	4120	Cleaned System II conc. compartments. Current density reduced in both systems.
3/8	63.3	38.2	3800	3720	
3/11	23.3	36.8	3090	3080	
3/12	70.0	46.4	3800	4260	
3/13	77.3	41.9	3965	4090	
3/20	53.3	35.0	3710	3415	Cleaned System I.

Table 9. ELECTRICAL RESISTANCE EXHAUSTION RUNS, Series 2 (Continued)

	5-cell resistance (ohms)		Dilute solution R_{ave} (ohm-cm)		
Date, 1963	System I	System II	System I	System II	Remarks
3/25	55.6	66.7	4390	5940	
3/26	46.9	54.8	5020	4260	
3/29	52.9	61.5	5590	5225	
4/1	18.8	26.5	2020	2170	
4/2	105.4	69.4	5650	5160	
4/3	102.8	63.9	5860	5320	
4/4	51.2	39.1	3530	3420	Added HCl to both conc. reservoirs.
4/6	114.3	55.0	-	-	
4/8	123.7	64.4	5860	4160	
4/9	132.3	81.6	5350	5400	
4/12	-	60.0	-	4360	Added HCl to both conc. reservoirs.
4/19	124.0	29.8	3975	3775	
4/23	-	48.9	-	4290	Added HCl to System I conc. Cleaned both systems.
4/25	27.7	24.2	4755	4830	
4/26	33.3	26.7	4440	4060	
4/29	60.9	21.7	5140	4350	Added HCl to both conc. reservoirs.
5/2	52.4	23.1	3700	3700	
5/9	27.6	20.0	2710	2835	Cleaned cathode System II.
5/10	52.4	23.3	2940	3480	24-hour conc. recycle begun.
5/11	52.9	30.0	4250	4255	
5/14	52.6	39.3	3280	2840	
5/17	53.5	28.9	3465	3590	
5/21	50.0	52.5	3630	3290	
5/22	65.7	35.7	4780	4140	
5/25	100.0	52.8	3565	3775	
5/28	60.4	49.0	2680	2650	

Because of calcium scaling and the fact that both systems were apparently operating polarized, electrical resistances were high in both systems until February 27. System II was disassembled on March 1 and $CaCO_3$ scaling was found in the concentrating compartments. This scale was dissolved with HCl, and the stack was reassembled. The diluting compartments were not disturbed. After reassembly, the 5-cell-pair resistance fell to about one-tenth its value when scaled. Thereafter, until April 4, this resistance fluctuated somewhat, but generally conformed to a pattern depending upon solution resistivity. From April 4 to April 24, when System II was again disassembled, the resistance increased as scaling developed and dropped again to the pattern of March 1 to April 4 as the scale was dissolved by acid treatment of the concentrate stream.

From March 1 to March 22, the electrical resistance of System I was, almost without exception, higher than that of System II. After System I was cleaned on March 22, its resistance initially fell below the pattern of System II. By April 2, however, scaling in the concentrate streams in System I was evidenced by the sharp drop in resistance after acid feed to the concentrate reservoir. Nevertheless, the resistance of System I remained consistently higher than that of System II.

On April 24, both stacks were disassembled. A noticeably greater amount of slime or organic deposition was found on the anion permeable membranes of System II than on those of System I. No scaling was found in System I, but the cathode compartment of System II was scaled heavily, and some light scaling was detected in the concentrating compartments. The heavier slime of System II was undoubtedly due to the fact that this system had been in operation, since its previous cleaning, approximately twice as long as had System I. Even with the heavier slime or organic deposition and the scaling of the concentrate streams, System II resistance, prior to cleaning, was considerably less than the resistance of System I.

After the final cleaning of the two systems on April 24, and until the end of the exhaustion series, the electrical resistance of System I remained consistently higher than that of System II. Since some suspended organic material was undoubtedly passing through the carbon adsorption column and depositing in System II, the consistently higher resistance exhibited by System I apparently was due not so much to this surface deposition or sliming, but to permeation and fouling of membrane interstices or pores by ABS or similar charged organic materials.

The erratic pattern of resistances of System I also indicates a building-up and periodic discharge of organic matter upon the membranes. Such a discharge, or sloughing, of organic deposits was frequently observed after shutting down and restarting System I. Occasional increases in COD, ABS and turbidity in the effluent stream also point to this conclusion. Similar sloughing was noted with System II although to a much lesser extent.

b. Current Efficiency

"Apparent" current efficiencies for the two systems in Series 2 were calculated (Table 10) according to the equation:

$$CE = \frac{(\Delta N,\text{meq/liter})\ (Q,\text{ml/min})}{(I,\text{ma})\ (3.106)} \qquad (1)$$

These current-efficiency calculations are based upon the difference between electrodialysis stack influent and effluent solution normality. The difference does not necessarily reflect the actual electrochemical species transferred across the membranes. Although influent and effluent bulk solution pH's are known, the pH in the boundary films and in the membranes themselves are not known. Consequently, carbonic species may enter and leave the electrodialysis stack as monovalent bicarbonate, but are transferred across the membranes as divalent carbonate. Similarly, the electrochemical valence of phosphate ions transferred across the membranes is probably not that of the phosphate ions in the bulk solution. Generally, an increase in influent solution pH has the effect of depressing the "apparent" current efficiency; that is, a current efficiency based upon the difference in influent and effluent solution normality. Nevertheless, it is the change in solution concentration that is of practical concern and so, with the reservations herein noted, the "apparent" current efficiency may be used as an additional index of electrodialysis stack performance.

Table 10 shows that the System II current efficiencies were nearly always lower than those of System I. There are several reasons why this might have occurred. System II could possibly have been electrically shorted, although examination of the stack revealed no obvious shorting problems. During most of the exhaustion runs, the effluent pH's from System II were higher than those from System I. Since even a small change in the pH of the bulk solution would indicate a large pH difference in the boundary films adjacent to the membranes, the differences noted in effluent pH's could be more significant than they at first appear. The higher boundary film and membrane pH's in System II would result in a lower "apparent" current efficiency because of the increased tendency to transfer carbonic and phosphate species as divalent rather than monovalent ions.

Bicarbonate removal is a particularly important factor because bicarbonate generally accounts for about half of the anion equivalents of the Clinton secondary effluent. The generally greater increase in hydronium ions (H_3O^+) in System I would result in an apparently greater bicarbonate removal, since some of the bicarbonate would be converted to dissolved carbon dioxide (H_2CO_3) according to the equilibrium:

$$HCO_3^- + H_3O^+ \rightleftharpoons H_2CO_3 + H_2O \rightleftharpoons CO_2 + 2H_2O \qquad (2)$$

Table 10. CURRENT EFFICIENCIES
EXHAUSTION RUNS, Series 2

	Current efficiency*		Effluent pH		
Date, 1963	System I	System II	System I	System II	Remarks
1/29	86.2	75.5	5.7	6.3	
2/5	89.9	64.4	6.3	6.9	
2/12	92.6	63.9	6.5	7.4	
2/19	63.8	59.6	6.9	7.4	System II cleaned.
2/27	72.6	108.2	7.1	7.3	
3/8	120.7	123.2	7.1	7.3	
3/12	113.1	89.6	7.2	7.5	
3/15	99.2	88.5	6.9	7.5	
3/19	119.9	93.8	7.4	7.4	System I cleaned.
3/26	98.3	73.8	6.7	6.8	
3/29	111.0	111.1	6.5	6.7	
4/2	115.2	64.9	6.6	6.7	
4/5	128.5	75.0	7.0	6.9	
4/9	128.2	90.3	7.5	7.4	
4/12	106.7	66.7	6.7	6.8	
4/19	166.9	92.1	6.8	6.8	
4/23	90.5	61.6	6.9	7.4	Both systems cleaned.
4/30	137.4	65.5	6.4	6.5	
5/3	95.3	-	6.8	6.9	System II scaling. 24-hour recycle begun.
5/10	-	31.9	-	6.5	
5/17	58.0	107.0	6.9	6.8	
5/24	93.1	64.8	7.0	7.0	
5/29	77.9	40.0	7.2	7.2	
5/31	-	59.0	-	7.4	

*Where complete cation analyses are available, current efficiencies are averages of those based upon anion and cation removals.

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An additional factor affecting current efficiency would be the ratio of product-to-waste stream concentrations. The effect of this parameter upon current efficiency is described in the section "Product-to-Waste Ratio Studies." Unfortunately, insufficient concentrate-stream data were obtained to evaluate the effects of the product-to-waste concentration ratio upon current efficiencies during the exhaustion series.

c. Energy Consumption

Energy consumptions for the two systems have been calculated (Table 11) according to the equation:

$$\text{Energy}\left(\frac{\text{watt-hrs}}{\text{eq removed}}\right) = 16.67 \frac{(V_5,\text{volts})\ (I,\text{ma})}{(Q,\text{ml/min})\ (\Delta N,\text{meq/liter})} \tag{3}$$

The energy consumption figures include the effects of both electrical resistance and current efficiency. Because of this, and because of the direct relationship between energy consumption and operating costs, the energy consumption figures offer a particularly applicable index to the performance of the electrodialysis test systems.

Table 11 shows that the energy consumptions for the two systems generally did not differ greatly throughout the exhaustion series. The higher electrical resistances of System I and the lower current efficiencies of System II apparently combined to give similar energy consumptions.

d. Limiting Current Density

In the idealized case of steady-state electrodialysis demineralization of a simple uni-univalent salt solution, the transfer of one ionic species (say, the anion across the anion permeable membrane) may be described by the relationship:[10]

$$D_L \frac{(C^\circ - C')}{\delta} = \frac{(t_m^- - t_s^-)\ i}{\mathcal{F}} \tag{4}$$

where:

D_L = diffusion coefficient of the salt (cm^2/sec);

C° = bulk solution concentration (eq/liter);

C' = solution concentration at membrane surface (eq/liter);

δ = boundary film thickness (cm);

t_m^- = transfer number of the anion in the membrane;

t_s^- = transfer number of the anion in the bulk solution;

Table 11. ENERGY CONSUMPTION EXHAUSTION RUNS, Series 2

Date, 1963	Energy consumption* $\left(\frac{\text{watt-hours}}{\text{eq removed}}\right)$ System I	System II	Remarks
1/29	75.4	-	
2/5	71.8	-	
2/12	78.8	94.1	
2/19	94.9	86.6	
2/27	21.4	41.3	
			System II cleaned.
3/8	8.4	5.7	
3/12	6.3	7.8	
3/15	9.2	-	
3/19	7.6	4.0	
			System I cleaned.
3/26	8.2	12.4	
3/29	8.7	11.6	
4/2	18.1	20.6	
4/5	16.7	15.7	
4/9	18.8	18.4	
4/12	-	19.3	
4/19	19.9	8.2	
4/23	-	28.7	
			Both systems cleaned.
4/30	10.9	12.3	
5/3	12.4	-	
			24-hour recycle begun.
5/10	-	17.5	
5/17	16.3	6.5	
5/24	17.5	23.1	
5/29	18.1	33.5	
5/30	-	-	

*Where complete cation analyses are available, energy consumptions are averages of those based upon anion and cation removals.

i = current density (ma/cm^2); and

$\mathcal{F}$ = Faraday's Constant.

Such a case is illustrated in Figure 9.

As the current density is increased, the transfer of the anion increases until its concentration at the membrane surface (C') approaches zero. By rearranging equation (4) and setting the surface concentration equal to zero, the standard limiting current density relationship is obtained.

$$\left(\frac{i}{C^\circ}\right)_{\lim} = \frac{\mathcal{F}\, D_L}{\delta(t_m^- - t_s^-)} \qquad (5)$$

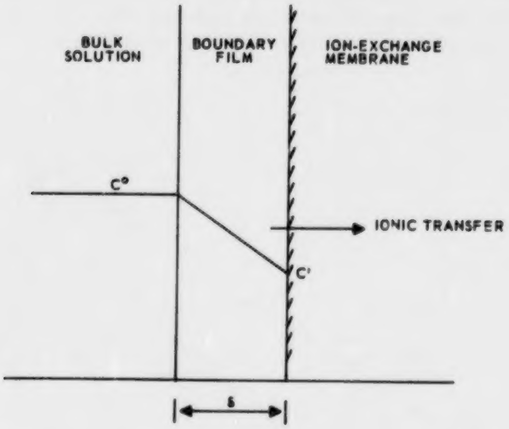


Figure 9. Concentration gradient across boundary (Nernst) film.

As the current density is increased beyond the limiting value, water is dissociated at the membrane surface (polarization) and the current density is increased by the transfer of hydroxyl ions (assuming polarization occurs at the anion membrane) across the membrane. The immediate net results of such a current density increase are the following:

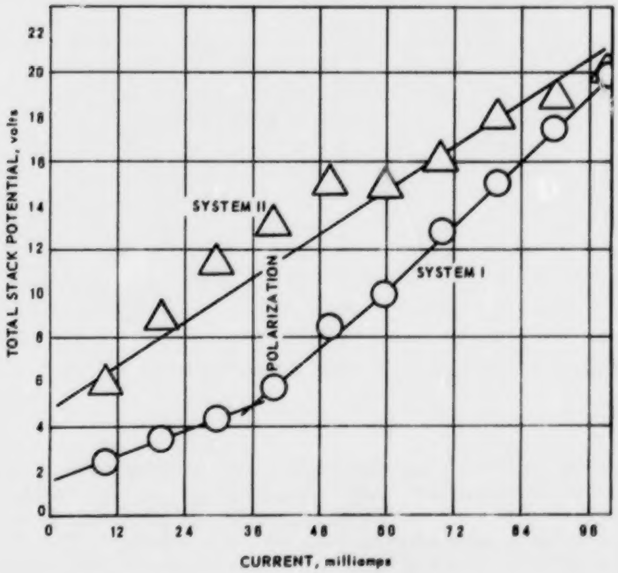
(1) a decrease in current efficiency, since the current carried by hydroxyl ions does not contribute to demineralization;

(2) an increase in electrical resistance due to the low film concentrations;

(3) an increase in energy consumption due to the combination of the above two results, and also because of the additional energy required for the dissociation of water;

(4) a change in pH at the membrane surface, which may result in the precipitation of pH sensitive components.

In actuality, limiting current density is not a point function, because competition between hydroxyl ions and other anions becomes significant as the film anion concentration approaches that of the hydroxyl ion. Also, in the case of a more complex solution, such as a secondary effluent, the limiting current density relationship is considerably more complex than is indicated by equation (5), the $D_L / (t_m^- - t_s^-)$ term being composed of some undefined combination of diffusion coefficients and transfer numbers. The dependency of the limiting current density upon the effective boundary film thickness, however, is very clear. The standard procedure of increasing the turbulence of the bulk solution to permit the use of higher current densities is based upon the decrease in the film thickness obtained from the increase in turbulence.



SYSTEM I: $(i/C)_{lim} = \frac{40}{212} \Big/ 0.00301 = 62.6 \frac{ma/cm^2}{eq/liter}$

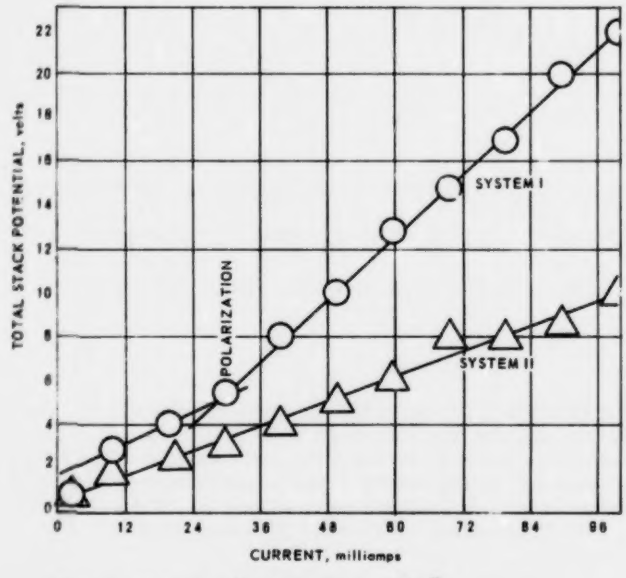
$v = 6.2$ cm/sec

Figure 10. Polarization runs I-1 and II-1 – February 19, 1963.

The effect of slime or organic deposits on the surface of the membranes is to increase the effective boundary film thickness, since its effective thickness then consists of the slime layer plus the major portion of the original film thickness. This increase in film thickness results in a lowering of the limiting current density, as indicated by equation (5). A reduction in the allowable current density means either a reduction in the amount of demineralization that can be accomplished with any specific electrodialysis stack, or operation in a polarized condition at greatly reduced efficiency. The limiting current density then becomes an important parameter of electrodialysis stack performance.

To determine the limiting current density for a particular solution at any one flow rate, we varied the stack voltage and plotted it against the resulting current density. A break in the slope of this curve occurs at the limiting current density.

The data for several polarization runs are plotted in Figures 10 through 13. The February 19 data (Figure 10) show System I beginning to polarize at an ($^{i}/C$) value of about 63 (see figure for units). The data for System II are somewhat erratic but no limiting ($^{i}/C$) is dis-

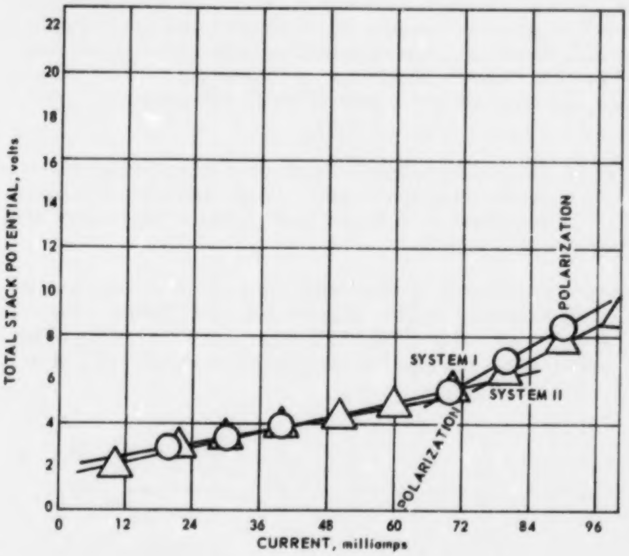


SYSTEM I: $(^{i}/C)_{lim} = \frac{30}{212} \Big/ 0.00306 = 46.2 \frac{ma/cm^2}{eq/liter}$

$v = 5.4$ cm/sec

Figure 11. Polarization runs I-2 and II-2 – March 18, 1963.

cernible in the range studied. The data indicate that the sliming of membranes in System I resulted in a reduction of the allowable curent density. The polarization runs conducted on March 18 (Figure 11) show polarization of System I beginning at an $(^{i}/C)$ value of about 46, and still no really discernible polarization point for System II.



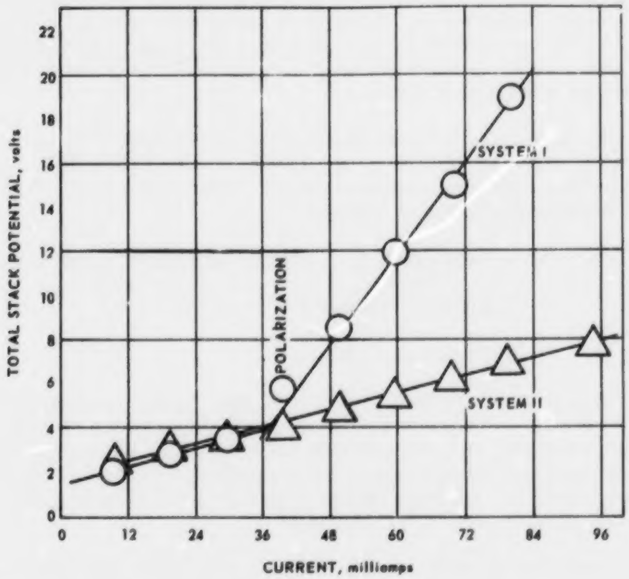
SYSTEM I: $(^{i}/C)_{lim} = \frac{70}{212} / 0.00262 = 126 \frac{ma/cm^2}{eq/liter}$ $v = 5.4$ cm/sec

SYSTEM II: $(^{i}/C)_{lim} = \frac{80}{212} / 0.00314 = 120 \frac{ma/cm^2}{eq/liter}$ $v = 6.45$ cm/sec

Figure 12. Polarization runs I-3 and II-3 – March 22, 1963.

On March 22 (Figure 12), after System I was disassembled and cleaned, the polarization point was increased to an $(^{i}/C)$ value of 126. The $(^{i}/C)_{lim}$ value for System II, on the same date, was 120. The final polarization run conducted in this series (April 18) is plotted in Figure 13. The $(^{i}/C)_{lim}$ value for System I had fallen to around 60. The polarization point for System II was again not found in the range studied.

The obvious deduction from these polarization data is that a relatively "clean" or organic-free influent to the electrodialysis stack is necessary to avoid either a greatly reduced allowable current density or frequent membrane maintenance.



$$\text{SYSTEM I: } ({}^{i}/C)_{1im} = \frac{40}{212} \Big/ 0.00327 = 57.7\, \frac{ma/cm^2}{eq/liter}$$

$$v = 6.7 \text{ cm/sec}$$

Figure 13. Polarization runs I-4 and II-4 – April 18, 1963.

e. Ionic Selectivity

The concept of selectivity is discussed at some length in Section III of this report, so it is described only briefly here. Generally, there is some preferential removal of one ionic species over others. This preferential removal, or selectivity, is a function of many factors. At one extreme, it is controlled by properties of the bulk solution; at the other extreme, it is controlled by properties of the membranes. In actual operation, the selectivities obtained will vary somewhere between the two extremes. Also, in actual operation, even the extremes may change. Solution properties may be varied; e.g., pH, valence, and temperature changes due to polarization. Membrane properties may be varied; e.g., surface characteristics and membrane capacity changes due to fouling. In addition, slime buildup or organic deposits at the membrane surfaces, and the subsequent increase in boundary film thicknesses may result in a closer approach to solution controlled selectivity.

In an electrodialysis system used in a waste-water renovation application, a change in selectivity is not necessarily detrimental; in fact, because its maximum allowable concentration is so low, a greatly increased preference for NH_4^+ removal may even allow such a system to be operated more efficiently.

A change in the relative rates of removal (selectivity) of the various ionic species, however, may be an indication of subtle changes in membrane properties that would not necessarily be reflected in other performance parameters so far discussed.

Selectivity may be described by a "separation factor" approximation:

$$\alpha^{i}_{T} \doteq \frac{\%\ \text{species i removed}}{\%\ \text{total meq removed}} \qquad (6)$$

As discussed in Section III of this report, this relationship is not strictly correct; however, the approximation is adequate for detecting trends, which is sufficient at this stage for estimating any major effects of secondary sewage effluent upon membrane selectivity. Separation factors calculated by equation (6) have been extracted from analytical data and are tabulated in Tables 12 and 13.

The general effects of the exhaustion runs upon ionic selectivity are summarized as follows:

Cations. It is difficult to detect any trends attributable to membrane fouling with the cation removal ratios. This is partly because separation factors appeared to be dependent upon the fraction total cut (which is a measure of approach to polarization and film diffusion limitations). Also, no particular fouling of cation membranes was experienced (the organic material, being primarily negatively charged, was attracted to the anion membranes). There is not sufficient data to charaterize sodium and potassium separations. Ammonium ions and total hardness (calcium and magnesium) were removed relatively faster in System II than in System I. While this could be indicative of membrane fouling, since System I was exposed to a higher organic concentration, a more likely reason is that separation factors for these ions appeared to decrease with an increasing total cut, the System II total cut being generally lower than that of System I.

Anions. The plots of chloride and nitrate removals (see Section III E) show that relative removals of both these species decreased with increasing total cut. The total cuts of System II were generally less than those of System I; therefore the chloride and nitrate separation factors from System II would be expected to be higher. The fact that the opposite was true indicates that anion membrane fouling increases chloride and nitrate separation factors. Conversely, the effect of anion membrane fouling upon bicarbonate removal was to depress the bicarbonate separation factor.

Table 12. EXHAUSTION RUNS - SYSTEM I

Separation Factors (α^i_T)

Date, 1963	Total cut %	i= Na$^+$	i= K$^+$	i= NH$_4^+$	i= Ca^{+2}+Mg^{+2}	i= Cl$^-$	i= HCO$_3^-$	i= NO$_3^-$	i= SO$_4^=$	i= PO$_4^\equiv$	Effl. pH	Remarks
1/29	76,6	0,95		1.09	0.68	1.00	1.16		0,67	0,65	5.7	
2/5	64.5	0.76	1.03	1,5	0,90	1.05	1,14		0.29	0,30	6,3	
2/12	62.5	0,92		0,80	0.64	1.23	0,95		0.83	0,93	6,5	
2/19												
2/27	13,8			2.14	1.56	1.98	0.75		1.12	1,26	7.1	
3/8	25,8	0.88	1.66	0,98	0.81	1.38	0.88		0.94	0.39	7,1	
3/12	19,3			1.29	1.22	1.61	0.52		0,74		7.2	
3/15	17,8			0,24	1.71	1.80	0,58		0,49	0,24	6.9	
3/19	21.9			1.17	1,43	1.52	0.89		0.74		7.4	System cleaned.
3/26	34.3			1.15	1.14	1.45	0.59	2.04	0.42	1.94	6.7	
3/29	38.7			0.93	0.98	1.28	0.88	1.85		0,65	6.5	

Table 12. (continued)

EXHAUSTION RUNS - SYSTEM I

Separation Factors ($\alpha \frac{i}{T}$)

Date, 1963	Total cut, %	i= Na^+	i= K^+	i= NH_4^+	i= $Ca^{+2}+Mg^{+2}$	i= Cl^-	i= HCO_3^-	i= NO_3^-	i= $SO_4^=$	i= $PO_4^\equiv$	Effl. pH	Remarks
4/2	46.4			0.81	1.24	1.25	0.89	1.45	0.41	1.23	6.6	
4/5	40.0			1.11	0.99	1.33	0.84		1.04	0.74	7.0	HCl to conc.
4/9	40.0				1.26	1.27	1.18	1.47	0.45	0.37	7.5	
4/12	31.7			1.90	1.08	1.43	1.08			0.81	6.7	HCl to conc.
4/19	34.7			1.02	1.05	1.47	0.60	1.77	0.86	0.33	6.8	HC1 to conc.
4/23	22.2			0.25	1.40	1.76	0.93				6.9	
4/30	41.3			2.06	1.17	1.39	0.87	1.61	0.42	0.30	6.4	System cleaned.
5/3	22.8			1.74	1.76	1.17	1.23		0.36	0.14	6.8	
5/17	18.5	1.45	0.71	-	1.89	2.05	0.57	2.13	0.60	0.83	6.9	24-hr. recycle started.
5/24	23.5	0.54	0.90	0.80	0.65	1.29	0.62	2.55	1.58	0.63	7.0	
5/29	19.2	0.45	0.76	0.69	2.31	2.08	0.68	2.60	0.87	0.35	7.2	

Table 13. EXHAUSTION RUNS - SYSTEM II*

Separation Factors (α_T^i)

Date, 1963	Total cut, %	i= Na^+	i= K^+	i= NH_4^+	i= $Ca^{+2}+Mg^{+2}$	i= Cl^-	i= HCO_3^-	i= NO_3^-	i= $SO_4^=$	i= $PO_4^{\equiv}$	Effl. pH	Remarks
1/29	57.5	1.22	1.26	1.05	1.20	0.99	1.13		1.10	0.68	6.3	
2/5	50.8	0.80	0.97	1.05	0.96	1.13	0.95		1.05	0.54	6.9	
2/12	31.7	0.99		1.14	1.24	1.25	0.79		1.50	1.89	7.4	
2/19	19.2	0.56	2.97	1.92	1.37	1.62	0.81		1.04	1.39	7.4	
2/27	15.5			1.10	0.64	0.92	0.55	2.38	0.61	2.79	7.3	System cleaned.
3/8	28.7	0.97	2.49	1.41	(-0.04)	1.13	0.96		0.10	1.50	7.3	
3/12	14.1			2.28	1.76	1.81	0.89		0.66	1.84	7.5	
3/15	16.3			-	1.75	1.24	0.90	1.60	0.83	0.31	7.5	
3/19	22.7			1.50	1.30	1.29	0.98		1.18	1.40	7.4	
3/26	22.5			1.83	1.28	1.40	0.83	1.16	0.67	1.33	6.8	
3/29	40.2			0.90	0.62	0.78	1.50	1.12		0.83	6.7	

*Values enclosed by parentheses are from periods of stack malfunctioning, i.e., severe scaling.

Table 13. (continued)

EXHAUSTION RUNS - SYSTEM II*

Separation Factors $(\alpha \frac{i}{T})$

Date, 1963	Total cut, %	i= Na^+	i= K^+	i= NH_4^+	i= $Ca^{+2} + Mg^{+2}$	i= Cl^-	i= HCO_3^-	i= NO_3^-	i= $SO_4^=$	i= $PO_4^{\equiv}$	Effl. pH	Remarks
4/2	22.1			1.48	1.63	1.52	1.26		0.13	2.71	6.7	
4/5	23.6			1.57	1.60	1.19	0.83	0.82	1.41	0.85	6.9	HCl to conc.
4/9	27.8			-	1.34	1.01	1.06	2.13	0.33	0.45	7.4	
4/12	22.2			2.68	1.45	1.37	1.20		-	0.64	6.8	HCl to conc.
4/19	25.4			1.22	1.24	1.14	0.97		1.38	0.13	6.8	
4/23	18.4			1.30	2.05	0.83	0.97	2.11	1.38	0.89	7.4	System cleaned.
4/30	37.3			1.08	1.17	(-0.14)	1.22	1.65	0.74	1.28	6.5	
5/10	5.7	2.26	2.56	(-2.26)	6.88	(-1.58)	3.33		(-10.32)	5.02	6.5	(Scaling) 24-hr recycle started.
5/17	33.8	1.03	0.89	1.09	1.37	0.82	0.52	2.74	1.03	0.95	6.8	
5/24	20.2	0.74	1.20	1.14	1.81	1.53	0.66		1.91	0.60	7.0	
5/29	11.1	0.58	0.90	0.28	3.62	1.35	0.85	3.00	0.86	0.75	7.2	
5/31	13.2			0.45	2.23	1.27	0.76		1.31	1.11	7.4	

*Values enclosed by parentheses are from periods of stack malfunctioning, i.e., severe scaling.

Sulfate removal was more difficult to characterize. Initially, in both systems, the sulfate separation factor appeared to increase (similar to that of bicarbonate) with increasing total cut. By the end of the runs in both systems, however, this dependency seemed to have reversed itself to the point where the sulfate separation factor began to decrease with increasing total cut. It would appear, therefore, that sulfate removal was somewhat enhanced by anion membrane fouling, but not to the extent of the increased chloride removal. This, incidentally, is in accord with certain field experience with demineralization of sulfate-containing well waters.

Phosphate separation factors were depressed by membrane fouling. These separation factors decreased with operation time in both systems; were increased temporarily after the systems were cleaned; and were consistently lower in the more fouled System I. These effects are illustrated clearly in the plots of Section III E.

f. Membrane Properties

At the conclusion of the product-to-waste ratio studies conducted after the second exhaustion series, the two electrodialysis stacks were disassembled and the membranes were subjected to Ionics' standard quality-control testing procedures. These test data were then compared to test data obtained prior to the exhaustion runs to determine whether any permanent changes in standard membrane properties had occurred. Data from these tests are tabulated in Tables 14 and 15.

Table 14. MEMBRANE ANALYSIS - SYSTEM I

August 5, 1963

Membrane properties	Membrane sample					
	CR-61AZL-183*			AR-111BZL-183*		
	1	2	Unused	3	4	Unused
Mullen burst strength, psi	90	95	100	105	105	
Thickness, cm	0.063	0.073	0.060	0.071	0.069	0.064
Resistivity, ohm-cm^2	10.1	9.8	11.8	66.4	68.3	13.5
Water content, %	55.9	55.6	57.6	41.0	41.2	41.0
Capacity, meq/dry gram	3.20	3.19	3.10	1.11	1.10	1.81

*Samples were taken from two cation and two anion membranes in the center of the stack.

Table 15. MEMBRANE ANALYSIS - SYSTEM II

August 5, 1963

Membrane properties	Membrane sample					
	CR-61AZL-183*			AR-111BZL-183*		
	1	2	Unused	3	4	Unused
Mullen burst strength, psi	105	105	100	110	115	
Thickness, cm	0.062	0.062	0.060	0.071	0.069	0.064
Resistivity, ohm-cm^2	9.54	9.71	11.8	15.5	17.2	13.5
Water content, %	56.8	57.1	57.6	45.2	45.0	41.0
Capacity, meq/dry gram	3.44	3.43	3.10	2.11	2.14	1.81

*Samples were taken from two cation and two anion membranes in the center of the stack.

The cation permeable membranes in both systems apparently did not deteriorate throughout the exhaustion runs and the product-to-waste ratio studies. The apparent increase in capacity is within the limits of expected sample variation. The anion membranes of System I, to the contrary, showed the effects of fouling as an increase in resistivity and a decrease in capacity. We suspected that the sorption of ABS or similar organic ions was responsible for this fouling. Additional samples of the used membranes were taken from both stacks and tested both before and after ABS elution with a 50/50 HC1/Acetone mixture. The results of these tests are summarized in Table 16. The membrane resistivities were restored very nearly to their original values by the elution of ABS. Capacities were not increased greatly. The ABS removal procedure consisted of four serial elutions of the 3/4- by 3-inch membrane samples in 100 milliliters of 1M 50/50 HC1/Acetone. Elution periods were 20 minutes each. A more prolonged procedure would possibly have removed more ABS, thereby restoring the membrane capacities nearer to their original values. Near quantitative ABS recoveries have been obtained previously from similar membranes by similar elution procedures. Alternatively, the remaining capacity loss possibly was due to other materials, such as organic acids, etc., that would require a different solvent for their extraction. This possibility was not investigated.

Table 16. ELUTION OF ABS FROM ANION MEMBRANES AR-111BZL-183

Sample properties	Unused membranes	SYSTEM I				SYSTEM II			
		Before elution		After elution		Before elution		After elution	
		1	2	1	2	1	2	1	2
Thickness, cm	0.064	0.067	0.067	0.065	0.066	0.065	0.065	0.071	0.066
Resistivity, ohm-cm^2	13.5	57.8	60.3	25.8	26.5	16.0	15.7	14.7	15.0
Water content, %	41.0	40.7	41.3	42.4	42.4	45.1	45.6	45.0	45.2
Capacity, meq/dry gram	1.81	1.11	1.13	1.19	1.18	1.69	1.69	1.64	1.64
ABS eluted, mg/cm^2				0.024	0.024			0.0083	0.0083

C. PRODUCT-TO-WASTE RATIO STUDIES

An important economic consideration in a waste-water reclamation system involves the ultimate disposal of the final waste stream. Consequently, it is desirable to obtain the most highly concentrated (least volume) waste stream consistent with efficient electrodialysis operation. The product-to-waste ratio studies were initiated to estimate this minimum ultimate disposal volume.

1. Apparatus

Both systems used in the two exhaustion series were rebuilt for the product-to-waste ratio studies. Higher capacity pumps were installed to permit operation at higher flow rates. Provision was made for continuous acid feed to both the concentrating and the cathode streams. Secondary sewage effluent was used as concentrate stream make-up instead of tap water as in the exhaustion series. Three additional Polyzinc/Rustoleum lined 275-gallon steel tanks for sewage storage were installed. Flow meters and pressure gauges were added to both systems. The activated carbon adsorption column in System II was enlarged to an 18-inch diameter column, 36 inches deep; this column was filled to half its depth, and upflow operation was initiated to obtain greater adsorption efficiencies.[11] A flow schematic of the rebuilt System II is shown in Figure 14. System I was identical except that it did not include the carbon column. The electrodialysis stacks in both systems were modified so that the electrode compartments would be diluting instead of concentrating streams. This modification was adopted to test the feasibility of using secondary sewage effluent for electrode streams; the electrodialysis stack configuration is illustrated in Figure 15. Other than the two additional membranes per stack which were required for this configuration, the membranes used in the two systems were those used in the second exhaustion series. Additional information on exhaustion was thereby obtained.

2. Test Procedure

System I was maintained at an approximate product-to-waste volume ratio of 10 to 1. This resulted in waste-to-product concentration ratios of less than 20 to 1. Dilute streamflow rates ranged from three to six times those used in the two exhaustion series. Initially, H_2SO_4 was employed as acid feed in the cathode and concentrating streams. Approximately 0.4 milliequivalent of acid feed per minute was used in each of these streams to maintain the pH's between 2 and 3. After the first week of operation, HCl was used instead of H_2SO_4, primarily because of precipitation difficulties incurred in System II.

System II was operated at varying flow rates and at steadily increasing product-to-waste volume ratios. At a 12.5 to 1 ratio, some scaling developed. This was not considered too meaningful, however, since the concentrate pH at this time was 6.4. Operation at higher product-to-waste volume ratios had been successful previously with lower concentrate pH's. The acid feed at this time had been converted from H_2SO_4

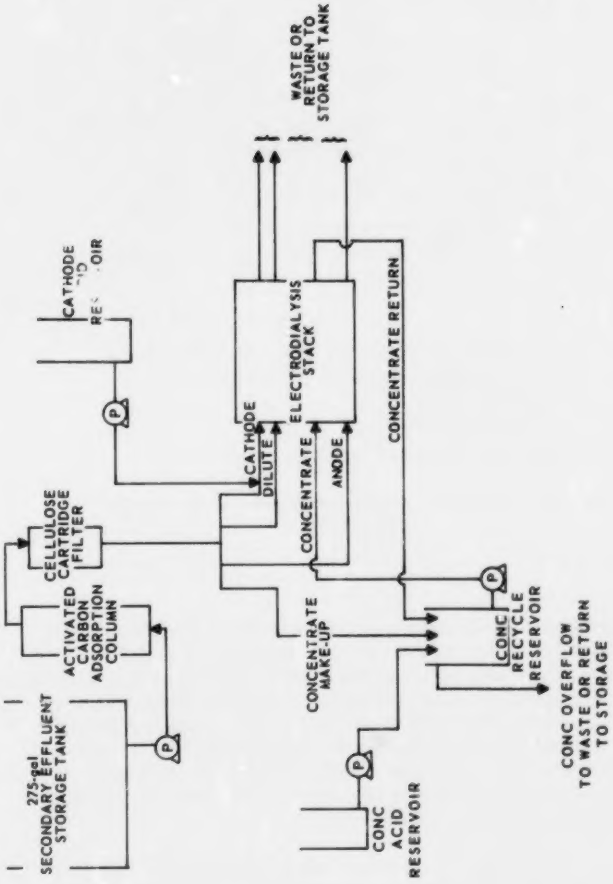
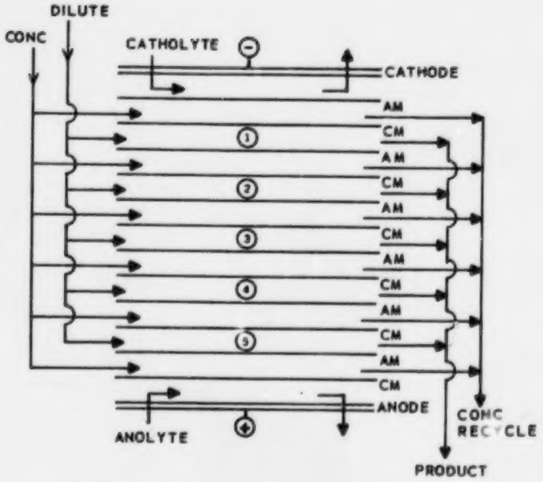

275-gal SECONDARY EFFLUENT STORAGE TANK
P
ACTIVATED CARBON ADSORPTION COLUMN
CELLULOSE CARTRIDGE FILTER
CATHODE
P
CATHODE
DILUTE
ELECTRODIALYSIS STACK
CONCENTRATE
ANODE
CONCENTRATE RETURN
WASTE OR RETURN TO STORAGE TANK
CONCENTRATE MAKE-UP
P
CONC RECYCLE RESERVOIR
P
CONC ACID RESERVOIR
CONC OVERFLOW TO WASTE OR RETURN TO STORAGE

Figure 14. Flow schematic, system II – product-to-waste ratio studies.



AM = ANION PERMEABLE MEMBRANE – AR-111BZL-183

CM = CATION PERMEABLE MEMBRANE – CR-61AZL-183

NO. OF CELL-PAIRS = 5

SPACER THICKNESS = 0.040 in.

EFFECTIVE MEMBRANE AREA = 212 cm^2 per cell-pair

Figure 15. Electrodialysis stack configuration – product-to-waste ratio studies.

to HCl. Cathode acid feeds between 0.3 and 0.8 milliequivalent per minute maintained the cathode stream pH between 1.5 and 3.0. Under these conditions, no scaling was encountered. An occasional acid pump failure at the high product-to-waste volume ratios resulted in very rapid scaling, particularly in the cathode compartment. Upon resumption of acid feed and subsequent lowering of pH, the scaling quickly stopped. During the final week of operation, the concentrate stream pH in System II was allowed to increase slightly in order to investigate selectivity dependence upon concentrate-stream pH.

At the end of the product-to-waste studies, both electrodialysis stacks were disassembled and the membranes tested for electrical resistance, capacity, and water content. These test results were reported in the discussion of exhaustion series 2.

Electrodialysis stack-operating parameters -- voltage, current, flowrate, pressure, solution pH's, temperatures, and resistivities -- were monitored throughout the product-to-waste studies. These data, as well as general operating information appear chronologically in Appendix A.

Voltage probes were conducted similar to those made during the two exhaustion series, the results of which are tabulated in Appendix C.

Demineralization information was obtained from analyses of influent- and effluent-stream grab samples; concentrate-stream samples were analyzed simultaneously to provide information for estimating waste-to-product concentration ratio effects upon selectivity of the system. These analyses were made twice weekly and consisted of: Na^+, K^+, NH_4^+, Ca^{+2}, Mg^{+2}, Cl^-, HCO_3^-, NO_3^-, $SO_4^=$, $PO_4^{\equiv}$, ABS, COD, organic nitrogen, alkalinity, pH, resistivity, temperature, turbidity, and gravimetric TDS determinations. The results are tabulated in Appendix B.

3. Test Results and Discussion

In the product-to-waste ratio studies, data were obtained on the same parameters of electrodialysis stack performance that were evaluated in the two exhaustion series. Since the membranes used in the exhaustion runs were not replaced for the product-to-waste ratio studies, additional information on membrane exhaustion was obtained; the main purpose of the studies, however, was the determination of the effect of increasing waste-to-product concentration ratios upon these performance parameters, and it therefore, received primary attention.

a. Electrical Resistance

Electrical resistance of an electrodialysis stack is affected by all the factors discussed in relation to the exhaustion series. In addition, an increase in concentration difference between diluting and concentrating compartments results in a greater tendency for back-diffusion (from concentrating into diluting compartments) and a higher voltage is required to overcome this tendency. This requirement is reflected as an increase in the electrical resistance of the system. The individual cell-pair resistances are tabluated in Appendix C, and resistances of the five cell-pairs are shown in Table 17.

A plot (Figure 16) of cell-pair resistance versus solution resistance for System II illustrates the general effect of concentration ratio and solution resistance upon the electrical resistance; the effect of membrane fouling or slime deposits is shown by the erratic resistance pattern of System I. This erraticism was also noticed throughout the exhaustion series and is interpreted as an indication of the buildup and periodic discharge or sloughing of organic deposits. Even when operating at very high waste-to-product concentration ratios, the electrical resistances of System II remained generally below those of System I. These factors indicate that, although increasing the concentration ratio results in a noticeably higher electrical resistance, the effect is much less severe than that of membrane fouling.

Table 17. ELECTRICAL RESISTANCE PRODUCT-TO-WASTE RATIO STUDIES

	Waste-to-product conc. ratio		5-cell-pair resistance		
Date, 1963	System I	System II	System I	System II	Remarks
6/26		11.0*		16.2	
6/28	8.1	4.9		134.0	
7/1	7.0	5.4	27.0	25.6	
7/3			27.3	30.9	
7/6			20.9	26.4	System II conc. makeup = 0. Cleaned both systems.
7/8		37.8	18.9	18.9	
7/9			7.0	7.4	
7/10			6.4	11.7	
7/11			14.2	13.6	
7/11				10.0	Added HCl to System II cathode stream.
7/12	10.9	39.2	10.6	7.5	
7/15			14.6	12.3	
7/16	15.1	46.6	15.0	13.1	
7/17			13.1	9.0	
7/18			23.0	12.9	
7/19	18.9	58.4	26.7	17.2	
7/22			24.3	24.7	
7/22				16.0	Cleaned both systems.
7/23	17.1	63.6	14.1	15.8	
7/24			28.9	20.6	
7/25			30.0	22.8	
7/26		75.5	43.3	35.4	
7/26			10.0	18.6	Added HCl to System II conc. stream.
7/27			28.0	30.3	
7/29		125.0	23.9	25.0	

*Estimated from incomplete analyses.

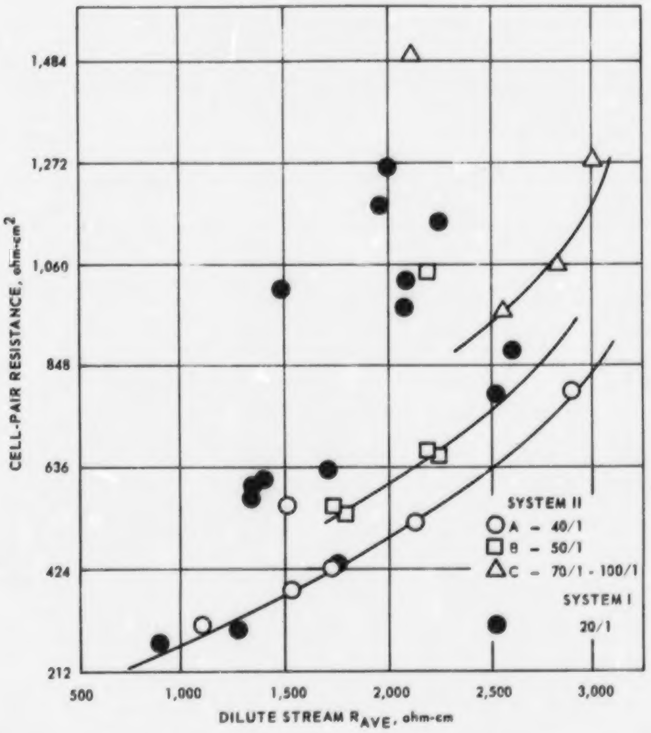


Figure 16. Electrodialysis stack resistance summary – product-to-waste concentration ratio, system II.

b. Current Efficiency

Current efficiencies were calculated from equation (1) and are tabulated in Table 18. As the concentration difference between diluting and concentrating compartments increases, the permselectivity of the membranes may be expected to decrease. Some anions may diffuse through the cation permeable membranes from the concentrating compartment back into the diluting compartment. Cations, likewise, may back-diffuse through the anion permeable membranes. The effect of this back-diffusion is a reduction in current efficiency of the system, since removal of some of the ions transported by electrical current is counteracted by those ions transported back by diffusion.

Table 18. PRODUCT-TO-WASTE RATIO STUDIES CURRENT EFFICIENCIES

	Waste-to-product concentration ratio		Current efficiency*		
Date, 1963	System I	System II	System I	System II	Remarks
6/26		11.0		63.8	
6/28	8.1	4.9	175.8	140.6	
7/2	7.0	5.4	162.1	92.9	
7/8		37.8		131.5	Cleaned both systems.
7/12	10.9	39.2	87.7	57.5	
7/16	15.1	46.6	126.4	72.2	
7/19	18.9	58.4	118.3	76.8	
7/23	17.1	63.6	83.2	52.8	Cleaned both systems.
7/26		75.5		67.7	
7/29		125.0	127.2	40.2	

*Current efficiencies are arithmetic averages of those calculated on an anion and a cation removal basis.

The current efficiencies of System II operating at increasing waste-to-product concentration ratios steadily decreased; moreover, they were consistently lower than the current efficiencies of System I, in which a concentration ratio of only about 10 or 20 was maintained. As further evidence of increased back-diffusion at the higher waste-to-product concentration ratios, a dependence of product-stream pH upon waste-stream pH in the absence of electrical current is shown in Table 19. This dependency was evidently the result of the back-diffusion of hydrogen ions rather than any polarization effects. The fact that pH decreases would have a tendency to increase the "apparent" bicarbonate removals and hence the "apparent" current efficiencies leads to the conclusion that true current efficiencies for System II were possibly lower than those indicated in Table 18.

Table 19. BACK DIFFUSION STUDY - SYSTEM II

July 18, 1963

Time	Current, ma	Waste pH	Product pH	Remarks
10:30 AM	160	1.8	6.6	
10:45 AM	0	1.8	6.7	
11:00 AM	-	-	-	Added NaOH to Conc.
11:20 AM	0	2.7	6.8	
11:25 AM	-	-	-	Added more NaOH.
11:30 AM	0	7.0	7.3	
11:35 AM	-	-	-	Added HCl to Conc.
11:40 AM	0	3.1	7.0	
12:50 PM	100	1.7	6.7	Normal operation.

Waste-to-product concentration ratio = 60/1

Table 20. PRODUCT-TO-WASTE RATIO STUDIES ENERGY CONSUMPTIONS

	Waste-to-product conc. ratio		Product-to-waste volume ratio		Energy consumption,* $\left(\frac{\text{watt-hr}}{\text{eq removed}}\right)$	
Date, 1963	System I	System II	System I	System II	System I	System II
6/26		11.0		8.3		57.1
6/28	8.1	4.9	7.2	8.9		13.1
7/2	7.0	5.4	9.6	8.7	10.2	17.3
7/8		37.8	23	330		7.3
7/12	10.9	39.2	10	-	10.7	17.1
7/16	15.1	46.6	10	250	7.7	15.7
7/19	18.9	58.4	9	300	15.4	30.3
7/23	17.1	63.6	11	250-300	30.1	65.9
7/26		75.5				67.3
7/29		125.0			19.8	107.0

*Energy consumptions are based on arithmetic averages of anion and cation removals.

c. Energy Consumption

Energy consumption is an important parameter for evaluation in the product-to-waste ratio studies. Since energy consumption is essentially an expression of operating cost, an optimized trade-off may be possible between ultimate waste-stream disposal cost and increased operating costs due to the lower waste-stream volumes.

Energy consumption values calculated from equation (3) are tabulated in Table 20. Concentration differences did not greatly influence energy consumption until a waste-to-product concentration ratio of about 50 was attained; beyond this value, energy consumption increased sharply. Whether the important parameter was the concentration ratio or the absolute concentration difference was not determined.

d. Limiting Current Density

During the exhaustion series, a relatively constant product-stream-flow rate was maintained. In the product-to-waste ratio studies, electrodialysis stack performance was investigated over a range of flow rates. Limiting current densities were determined by polarization runs conducted under various operating conditions. Two methods were used: plotting current density against a varying stack voltage at constant flow rate, and plotting current density against a varying flow rate at constant stack voltage. In both methods, the limiting current density was indicated by a change in the slope of the resulting curve at the "polarization point" of the system.

As shown by equation (5), the limiting current density for any electrodialysis application is inversely proportional to the boundary film thickness that is a function of flow path velocity. When no sliming or other surface fouling occurs this relationship can be described as:[10]

$$\delta = 0.54 \left(\frac{D_L \; D_e \; L}{V} \right)^{0.33} \tag{7}$$

where:

D_e = equivalent hydraulic diameter (cm);

L = flow path length between turbulence-promoting spacer webs (cm);

V = average linear flow path velocity (cm/sec).

Data extracted from the various polarization runs are tabulated in Tables 21 and 22; the function (^{1}/C) is calculated from:

$$(i/C) = \frac{I \, \rho \text{ ave}}{2{,}015} \qquad (8)$$

where:

I = stack current (ma);

ρ ave = average 25°C solution resistivity (ohm-cm).

Table 21. POLARIZATION DATA - SYSTEM I

Date, 1963	I_{lim}, ma	v, cm/sec	ρ_i,[a] ohm-cm	ρ_e,[a] ohm-cm	$(i/C)_{lim}$	Remarks
6/21	300	19.1	3,000	4,400	550	Initial Test Runs.
	475	27.3	3,000	4,300	860	
	170	8.25	3,000	4,800	329	
	380	21.6	3,000	4,400	700	
7/1	135	9.8	3,200	5,000	275	
7/10	605	25.8	1,000	1,500	375	
7/22	385	37.1	1,800	2,500	412	Stack fouling.

[a] Estimated.

Equation (8) was derived from the following:

$$(i/C) = \frac{(i,\ \text{ma/cm}^2)}{(C,\ \text{eq/liter})} \qquad (9)$$

$$= \frac{(I,\ \text{ma})}{(212\ \text{cm}^2)\ (C,\text{meq/liter})} \left(1{,}000\ \frac{\text{meq}}{\text{eq}}\right)$$

Table 22. POLARIZATION DATA - SYSTEM II

Date, 1963	I_{lim}, ma	v, cm/sec	ρ_i, ohm-cm	ρ_e, ohm-cm	$(i/C)_{lim}$	Remarks
6/28	160	12.9	3,000	5,500	338	Stack beginning to scale.
	180	18.0	3,000	5,000	358	
	265	28.3	3,000	4,800	514	
7/1	207	12.4	3,200	5,000	422	Stack clean.
7/10	830	23.2	1,000	1,700	556	
7/11	215	17.3	1,600	1,800	182	Stack scaled.
	405	26.8	1,600	1,700	332	
7/22	> 560	37.9	2,000	~ 3,400	> 750	

Based upon log-log plots of concentration versus 25°C resistivity, with a temperature correction of 2 percent per degree centigrade,

$$\rho_{25^\circ} = \rho_T \left[1 - 0.02\,(25-T)\right] , \qquad (10)$$

the general relationship between concentration and solution resistivity was:

$$C,\ \text{meq/liter} = \frac{9{,}500}{\rho_{25^\circ},\ \text{ohm-cm}} \qquad (11)$$

Log-log plots in Appendix D show this relationship. Serial dilutions and resistivity measurements made on the analyzed influent, effluent, and concentrating streams from System II on July 8, 1963, are plotted in Appendix D, the plot also follows the relationship described by equation (11).

By combining equations (9) and (11), the current density function was obtained:

$$({}^{i}/C) = \frac{(I,\text{ma})}{(212\ \text{cm}^2)}\ \frac{\rho_{25^\circ}}{(9500)}\ \left(1{,}000\ \frac{\text{meq}}{\text{eq}}\right)$$

$$({}^{i}/C) = \frac{I\ \rho_{25^\circ}}{2{,}015} \qquad (8)$$

The ρ_{25° used in calculations was an average of influent and effluent values of ρ_{25° and noted by ρ_{ave}.

The limiting current density function, $(i/C)_{lim}$, is plotted against flow path velocity in Figure 17, which shows that fouling and/or sliming of membrane surfaces lower the allowable $(i/C)_{lim}$, because of increased film thicknesses. Curve A is an envelope enclosing all data points and describes a clean system, prior to any fouling or slime buildup; curves B and C simply indicate the effects of increased fouling and scaling of membrane surfaces.

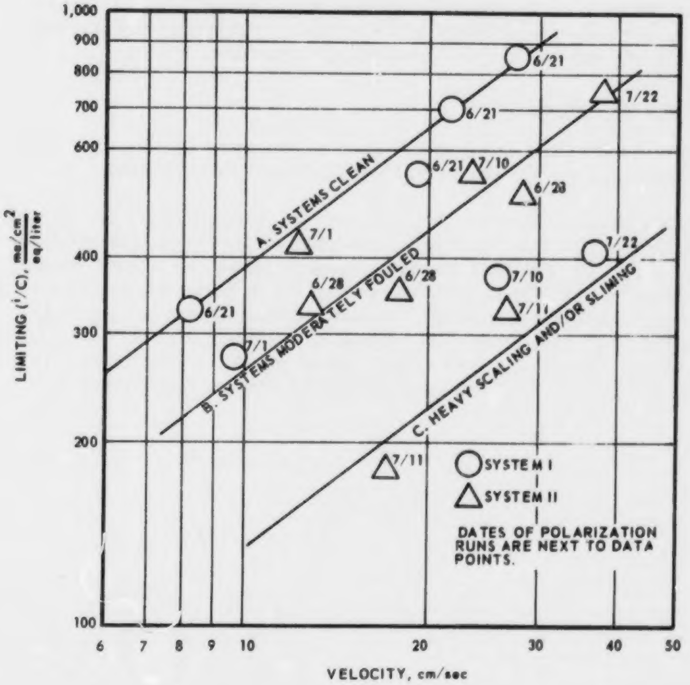


Figure 17. Limiting current density curves.

For a simple uni-univalent salt solution, equations (5) and (7) can be combined and rearranged to yield the theoretical relationship:

$$(i/C)_{lim} = \frac{\mathcal{F}}{0.54} \cdot \left(\frac{D_L^{2/3}}{t_m - t_s}\right)\left(\frac{V}{D_e L}\right)^m \qquad (12)$$

Experimental results with a sodium chloride solution, at 25°C, agree fairly well with this equation,[10] m being equal to 0.33. The empirical curve A of Figure 17 shows an m nearer to 0.7. Mason and Kirkham[12] used an m equal to 0.6.

Interestingly, this relationship is comprised of two distinct parameters: a diffusion parameter, controlled by properties of the solution and membranes; and a hydraulic parameter, controlled by the interior geometry of the electrodialysis stack and the flow path velocity. Achieving the highest degree of demineralization with any given electrodialysis stack requires operation at the highest allowable ($^{i}/C$) value. Equation (12) shows there are two separate approaches to increasing this allowable value: through manipulation of the system hydraulics; and through manipulation of solution and/or membrane properties.

e. Ionic Selectivity

Separation factors, described by equation (6), were calculated for the various ionic species during the product-to-waste ratio studies.

In Section III E of this report, individual species removals are plotted against total cut, and the effect of concentration ratio upon selectivity is discussed. Only the chloride and bicarbonate separation factors showed distinct changes from those calculated from the exhaustion runs. Chloride α decreased due to the use of HC1 as concentrate stream acid feed and the resulting disproportionate increase in concentrate stream chloride concentration. The bicarbonate α increased due to the conversion of bicarbonate to CO_2 in the acid condition of the concentrate streams.

SECTION III
SELECTIVITY

A. MECHANISM OF ION TRANSPORT

Before any discussion of selectivity in an electrodialysis application, an understanding of the mechanisms involved in the transport of ionic species across ion-exchange membranes and their complex interactions is necessary. The transport of an ionic species is described as a flux, J, which is the amount of the species passing through 1 square centimeter of membrane per second. The overall flux of the arbitrary species i, is composed of three additive terms: the diffusion flux $(J_i)_{diff}$ caused by the chemical potential gradient of the species; the electric transference $(J_i)_{elec}$ caused by the electrical potential gradient; and the transfer $(J_i)_{con}$ caused by convection.[13,14,15]

The diffusion flux is:

$$(J_i)_{diff} = - \mathcal{U}_i C_i \operatorname{grad} \mu_i \tag{13}$$

where:

$\mathcal{U}_i$ = diffusion mobility of species i ($cm^2 sec^{-1} joule^{-1} mole$);

C_i = concentration of species i (moles cm^{-3});

μ_i = chemical potential of species i.

since:

$$\mu_i = \mu^\circ_i + RT \ln C_i f_i \tag{14}$$

where:

μ_i° = chemical potential in the standard state;

R = universal gas constant;

T = absolute temperature;

f_i = molar activity coefficient of species i;

the gradient of the chemical potential of species i is:

$$\left.\begin{aligned} \operatorname{grad} \mu_i &= RT \operatorname{grad} (\ln C_i f_i) \\ &\text{or} \\ \operatorname{grad} \mu_i &= \frac{RT}{C_i} \operatorname{grad} C_i + RT \operatorname{grad} (\ln f_i) \end{aligned}\right) \tag{15}$$

The diffusion flux may thus be expressed as:

$$(J_i)_{diff} \quad = \quad - \mathcal{U}_i C_i RT \text{ grad } (\ln C_i f_i) \qquad (16a)$$

or

$$(J_i)_{diff} \quad = \quad - \mathcal{U}_i RT \left\{ \text{grad } C_i + C_i \text{ grad } (\ln f_i) \right\}. \qquad (16b)$$

The electrical transference is:

$$(J_i)_{elec} \quad = \quad - \mathcal{U}_i C_i \mathcal{F} Z_i \text{ grad } \phi \qquad (17)$$

where:

$\mathcal{F}$ = Faraday's Constant;

Z_i = electrochemical valence of species i;

ϕ = electrical potential (volts).

The transfer by convection is:

$$(J_i)_{con} = \bar{C}_i \sigma \qquad (18)$$

where:

σ = linear rate of motion of the pore liquid within the membrane (cm/sec);

$\bar{C}_i$ = concentration of species i within the membrane phase.

Moreover:

$$\sigma \quad = \quad \frac{\omega \mathcal{F} \bar{X}}{\rho_\circ \varepsilon} \text{ grad } \phi = \omega \bar{U}_\circ \text{ grad } \phi \qquad (19)$$

$$\bar{U}_\circ \quad = \quad \frac{\mathcal{F} \bar{X}}{\rho_\circ \varepsilon})$$

where:

ω = electrochemical valence of fixed ionic groups;

$\bar{X}$ = concentration of fixed ionic groups;

ε = fractional pore volume of membrane;

ρ = specific flow resistance of the membrane;

$\bar{U}_o$ = "mobility" of pore liquid within the membrane.

Only the convection transference $(J_i)_{con}$ specifically refers to the membrane phase; the diffusion flux $(J_i)_{diff}$ and electrical transference $(J_i)_{elec}$ relationships are sufficiently general to apply to either solution or membrane phase.

With the added definition:

$$\mathcal{U}_i = \frac{U_i}{|Z_i|\mathcal{F}} \tag{20}$$

where:

U_i = electrical mobility of species i ($cm^2 sec^{-1} volt^{-1}$);

$|Z_i|$ = absolute valence of species i.

The overall flux of an arbitrary species i in the membrane phase may be described:

$$J_i = \frac{-\bar{U}_i \bar{C}_i}{|Z_i|\mathcal{F}} \left[RT \text{ grad } (\ln \bar{C}_i \bar{\gamma}_i) + Z_i \mathcal{F} \text{ grad } \phi \right] + \omega \bar{U}_o \bar{C}_i \text{ grad } \phi \tag{21}$$

The notation $\bar{U}_i$, $\bar{C}_i$, $\bar{\gamma}_i$, refers to the mobility, concentration, and activity coefficient of species i in the membrane phase. The flux equation (21) applies to all mobile species in the system. The set of flux equations, one equation for each species, is subject to the restriction of electroneutrality:

$$\sum_i = Z_i \bar{C}_i + \omega \bar{X} = 0 \tag{22}$$

The set of flux equations, moreover, is related to the current density (assuming 100% current efficiency) by:

$$I/_A = \mathcal{F} \sum_i Z_i \quad J_i \tag{23}$$

The flux equations, the electroneutrality, and current density relationships for an anion permeable membrane used in the electrodialysis demineralization of a secondary sewage effluent are shown in Table 23.

Table 23. FLUX EQUATIONS FOR AN ANION PERMEABLE MEMBRANE

Equation No.	Equation	Notation	Species
(T1)	$J_i = \frac{-\bar{U}_i \bar{C}_i}{\mathcal{F}} \left[RT \text{ grad } (\ln \bar{C}_i \bar{f}_i) - \mathcal{F} \text{ grad } \phi\right] + \omega \bar{U}_o \bar{C}_i \text{ grad } \phi$	i	Cl^-
(T2)	$J_j = \frac{-\bar{U}_j \bar{C}_j}{\mathcal{F}} \left[RT \text{ grad } (\ln \bar{C}_j \bar{f}_j) - \mathcal{F} \text{ grad } \phi\right] + \omega \bar{U}_o \bar{C}_j \text{ grad } \phi$	j	HCO_3^-
(T3)	$J_k = \frac{-\bar{U}_k \bar{C}_k}{\mathcal{F}} \left[RT \text{ grad } (\ln \bar{C}_k \bar{f}_{k)} - 2\mathcal{F} \text{ grad } \phi\right] + \omega \bar{U}_o \bar{C}_k \text{ grad } \phi$	k	$CO_3^=$
(T4)	$J_l = \frac{-\bar{U}_l \bar{C}_l}{\mathcal{F}} \left[RT \text{ grad } (\ln \bar{C}_l \bar{f}_l) - \mathcal{F} \text{ grad } \phi\right] + \omega \bar{U}_o \bar{C}_l \text{ grad } \phi$	l	NO_3^-
(T5)	$J_m = \frac{-\bar{U}_m \bar{C}_m}{\mathcal{F}} \left[RT \text{ grad } (\ln \bar{C}_m \bar{f}_m) - 2\mathcal{F} \text{ grad } \phi\right] + \omega \bar{U}_o \bar{C}_m \text{ grad } \phi$	m	$SO_4^=$
(T6)	$J_n = \frac{-\bar{U}_n \bar{C}_n}{\mathcal{F}} \left[RT \text{ grad } (\ln \bar{C}_n \bar{f}_n) - \mathcal{F} \text{ grad } \phi\right] + \omega \bar{U}_o \bar{C}_n \text{ grad } \phi$	n	$H_2PO_4^-$
(T7)	$J_p = \frac{-\bar{U}_p \bar{C}_p}{\mathcal{F}} \left[RT \text{ grad } (\ln \bar{C}_p \bar{f}_p) - 2\mathcal{F} \text{ grad } \phi\right] + \omega \bar{U}_o \bar{C}_p \text{ grad } \phi$	p	$HPO_4^=$

GPO 802-360-3

Table 23. (continued)

FLUX EQUATIONS FOR AN ANION PERMEABLE MEMBRANE

Equation No.	Equation	Notation	Species
(T8)	$J_q \quad \frac{-\bar{U}_q \bar{C}_q}{\mathcal{F}} \left[RT \text{ grad } (\ln \bar{C}_q \bar{f}_q) - 3\mathcal{F} \text{ grad } \phi\right] + \omega U_o C_q \text{ grad } \phi$	q	$PO_4^{\equiv}$
(T9)	$J_r = \frac{-\bar{U}_r \bar{C}_r}{\mathcal{F}} \left[RT \text{ grad } (\ln \bar{C}_r \bar{f}_r) - \mathcal{F} \text{ grad } \phi\right] + \omega U_o C_r \text{ grad } \phi$	r	OH^-
(T10)	$- \bar{C}_i - \bar{C}_j - 2\bar{C}_k - \bar{C}_l - 2\bar{C}_m - \bar{C}_n - 2\bar{C}_p - 3\bar{C}_q - \bar{C}_r + \omega X = 0$		
(T11)	$\mathcal{F} \; (J_i + J_j + 2J_k + J_l + 2J_m + J_n + 2J_p + 3J_q + J_r \; [\; = I/A$		

These equations must be solved simultaneously to arrive at the flux J_i of any one species.

This is, in fact, a greatly simplified treatment. The most important simplifications are:

1. An idealized perm-selective membrane has been assumed; i.e., one with a fixed ion concentration that is so high that co-ions (in this case cations) are completely excluded. As waste-to-product concentration ratios increase, this becomes an increasingly poor assumption; flux equations, therefore, should be included for all co-ion species as well as for co-ion concentrations in the electroneutrality relationship.

2. Steady state conditions are assumed; i.e.:

$$\text{div } J_i = 0 \tag{24}$$

3. The flux equations, in the form given here, imply the validity of the diffusion and electrical mobility relationship (equation (20)). This essentially is equivalent to neglecting "coupling" of individual fluxes by other than electrical forces and convection.

4. Any differences in interaction of the various species with structural parts of the membrane have been neglected in the convection relationship (equation (18)).

5. It is also assumed that species cross the membrane as such (no complexing with other species within the membrane) and that their motion is not geometrically coupled (as in single-file diffusion).

6. Complete membrane diffusion control has been assumed. A somewhat similar set of flux equations might have been derived had complete film diffusion control been assumed. Such a system is considerably more complex, however, since all mobile species, anions and cations, must be considered. Moreover, the set of resulting equations must be solved simultaneously for both films and the membrane.

In practice, ionic transference by electrodialysis is neither completely membrane-diffusion nor film-diffusion controlled, but lies somewhere between.

With the simplifications noted, equations (21), (22), and (23), constitute a mathematical basis for the quantitative treatment of selectivity in an electrodialysis system. With certain further simplifications, general solutions to similar sets of equations have been obtained,[13] but they are exceedingly complex and difficult to evaluate.

The rate of removal (flux) of any particular species is dependent upon the rates of removal of all other species and, in this manner, the rate becomes a function not only of the particular species' concentration and mobility, but also of all other species' concentrations and mobilities. In an attempt to obtain usable information regarding the rates of removal of the various ionic species present in a secondary sewage effluent, without becoming embroiled in unjustifiable mathematics, three separate approaches have been used:

1. Some effects of certain external solution parameters have been examined by studying relative transfer rates across an anion permeable membrane;

2. possible effects of certain membrane parameters have been determined by estimating relative transfer rates from membrane equilibrium data; and

3. the overall effects of various operating conditions, including membrane fouling and scaling, have been estimated from information collected throughout the exhaustion series.

B. SEPARATION FACTOR

The concept of membrane selectivity involves the ability of an ion-exchange membrane to transfer ionic species preferentially. This selectivity may be described, for most practical purposes, by a separation factor, α:

$$\alpha_k^i = \frac{d\,C_i}{C_i} \cdot \frac{C_k}{d\,C_k} \qquad (25)$$

where:

α_k^i = the relative separation of species i over species k.

This expression obviously may also be written:

$$\alpha_k^i = \frac{J_i / C_i}{J_k / C_k} \qquad (26)$$

By the above definitions, a completely non-selective membrane would have a separation factor equal to unity for all combinations of species, i.e., the overall flux of all species would be identically proportional to the external solution concentration of that species:

$$\frac{J_i}{C_i} = \frac{J_i}{C_j} = \frac{J_k}{C_k} = \frac{J_1}{C_1} \; \ldots\ldots = \frac{J_n}{C_n} \qquad (27)$$

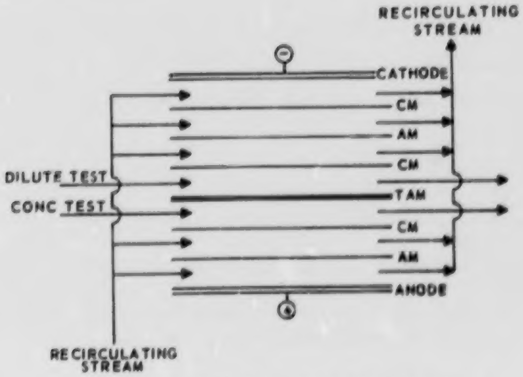
In view of flux equation (21), the theoretical significance of the separation factor defined by equations (6), (25), (23) is questioned. This separation factor is quite useful, however, in estimating the overall effects of various solution and operating parameters upon relative rates of removal. It is, moreover, easily adapted to engineering calculations to estimate removal rates and to determine limiting ionic species for design purposes.

C. EFFECTS OF SOLUTION PARAMETERS UPON A SEPARATION FACTOR

Field experiences in electrodialysis and general fundamentals of ion-exchange [13,14] have shown that separation factors are dependent upon such solution parameters as: total solution concentration, solution pH and temperature, fractional concentration of the species of concern, and the nature and fractional concentration of competing species. The flux equation (21) shows such a dependency, the individual membrane concentrations $\bar{C}_i$ and valences Z_i being related to external solution characteristics. In this study, the effects of pH and fractional concentration were examined in a phosphate - chloride system.

1. Apparatus

The electrodialysis stack used in this study was similar to those used in the exhaustion runs. The stack configuration (System III) is illustrated in Figure 18. This stack was modified to permit pumping separately to an isolated test cell-pair, which consisted of a diluting stream and a concentrating stream separated by an anion permeable membrane. Selective transport was observed by the species concentration changes across this test membrane.



CM = STANDARD CR-61AZL-183 MEMBRANE

AM = STANDARD AR-111BZL-183 MEMBRANE

TAM = ANION MEMBRANE BEING TESTED (AR111BZL-183)

Figure 18. Electrodialysis stack configuration, system III - selectivity studies.

2. Test Procedure

A potassium chloride - potassium phosphate solution was pumped through the dilute test compartment. The total concentration of chloride plus phosphate was maintained at approximately 5 millimoles per liter, but the ratio of chloride to phosphate was varied. Dilute-stream pH was varied by addition of KOH or H_3PO_4. A 5 millimoles-per-liter potassium nitrate solution was pumped through the concentrate test compartment. To supply electrical conductivity through the stack, a 5-millimoles-per-liter KNO_3 solution was recirculated through all other compartments. Flow-path velocities were maintained at about 6.4 centimeters per second. Stack current was maintained at 20 milliamps. With this particular configuration, nitrate was diffusing against the current from the concentrating test compartment into the diluting test compartment. The actual ionic system described, then, is a phosphate - chloride system in the presence of nitrate.

During each run, influent-and effluent-test-stream samples were analyzed, after a 2-nour equilibration period, for chloride, phosphate, and nitrate. Separation factor calculations are tabulated in Table 24.

3. Test Results and Discussion

Figure 19 illustrates clearly the effects of both pH and fractional concentration of phosphates upon the phosphate - chloride separation factor. At pH 4.5, when phosphate is predominantly monovalent $H_2PO_4^-$, this separation factor lies between 0.4 and 0.5 and increases only slightly with decreasing phosphate fraction. At pH 8.5, when phosphate is pre-

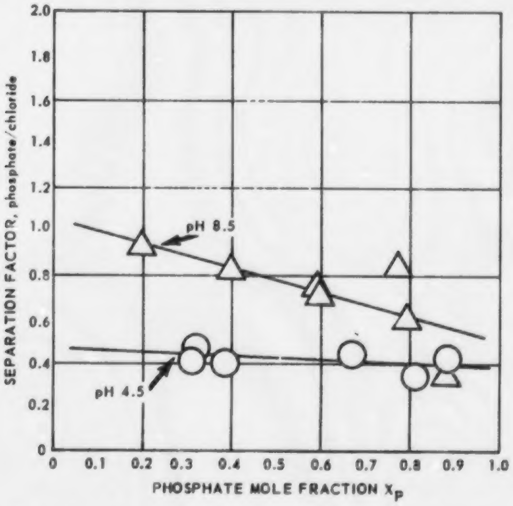


Figure 19. Calculated separation factors – phosphate/chloride system.

Table 24. EFFECT OF FRACTIONAL CONCENTRATION AND pH UPON α_{Cl}^{P}

Run No.	Phosphate			Chloride			Dilute pH_{ave}	X_p	α_{Cl}^{P}
	$(DI-DE)Q_D$	$(CE-CI)Q_C$	$\sqrt{DI \cdot DE}$*	$(DI-DE)Q_D$	$(CE-CI)Q_C$	$\sqrt{DI \cdot DE}$*			
A-2a	7.7	4.5	2.20	27.5	27.6	4.64	(est.	0.32	0.47
b	6.4	4.8	2.23	26.2	29.1	4.66	4.5)	0.32	0.42
A-3	9.7	6.2	1.90	31.5	31.6	3.01	(est. 4.5)	0.39	0.40
A-4	22.2	14.6	3.46	16.7	22.4	1.62	4.5	0.68	0.44
A-5	19.8	21.4	4.69	10.2	16.3	1.05	4.5	0.82	0.35
A-6	40.0	24.7	5.68	7.4	10.6	0.68	4.4	0.89	0.43
B-1	3.0	8.4	0.95	22.9	21.0	3.50	7.9	0.21	0.95
B-2	7.3	10.9	1.87	15.4	15.6	2.69	8.3	0.41	0.84
B-3	8.6	13.2	2.95	8.9	10.1	1.89	8.4	0.61	0.73
B-4	5.7	14.6	4.06	2.7	5.6	1.02	8.8	0.80	0.61
C-1	9.0	10.5	2.86	7.0	8.9	1.79	8.4	0.61	0.76
C-2	14.5	13.4	3.70	4.0	5.3	1.06	8.6	0.78	0.87
C-3	13.2	13.7	4.26	7.2	3.1	0.59	9.6	0.88	0.36

*Geometric mean values are used for average concentrations because concentrations change exponetially along the flow path length during electrodialysis of dilute solutions.

Table 24. (continued)

EFFECT OF FRACTIONAL CONCENTRATION AND pH UPON α^{P}_{Cl}

Run No.	Phosphate			Chloride					
	$(DI-DE)Q_D$	$(CE-CI)Q_C$	$\sqrt{DI \cdot DE}$*	$(DI-DE)Q_D$	$(CE-CI)Q_C$	$\sqrt{DI \cdot DE}$*	Dilute pH_{ave}	X_p	α^{P}_{Cl}
D-1	0	0.5	0.95	5.9	7.0	3.96	12.5	0.19	0.16
D-2	0	0.3	1.90	1.6	5.6	3.02	12.4	0.39	0.07
D-3	3.0	0	2.90	0	2.8	1.95	12.4	0.60	0.72
D-4	4.3	0	3.86	0	1.7	0.99	12.5	0.80	0.66

Notation:

DI = Concentration of dilute test influent
DE = Concentration of dilute test effluent
CI = Concentration of concentrate test influent
CE = Concentration of concentrate test effluent
Q_D = Dilute test flow rate
Q_C = Conc. test flow rate
X_P = Phosphate mole fraction

$$X_P = \frac{(DI \cdot DE)^{1/2}_{P}}{(DI \cdot DE)^{1/2}_{P} + (DI \cdot DE)^{1/2}_{Cl}}$$

*Geometric mean values are used for average concentrations because concentrations change exponetially along the flow path length during electrodialysis of dilute solutions.

dominantly divalent $HPO_4^=$, the separation factor is not only greater (0.6 to 1.0), but increases at a higher rate with decreasing phosphate fraction. The data taken at pH 12.5 are not plotted because the overwhelming OH^- concentrations made these data difficult to evaluate.

The results of this study indicate that the relative rate of phosphate removal can be increased by raising the solution pH, and that phosphate will move faster than chloride, through the anion membrane tested, only at very low (<0.1) fractional phosphate concentrations and at high pH's (>8.0).

D. EFFECTS OF MEMBRANE PARAMETERS UPON A SEPARATION FACTOR

Examination of flux equation (21) shows the rate of transfer of any species i to be proportional to the product of its membrane concentration $\bar{C}_i$, its mobility in the membrane phase $\bar{U}_i$, and the total driving force acting upon the species (assuming complete membrane control and ignoring convection):

$$J_i \doteq \frac{-\bar{U}_i \bar{C}_i}{|z| \mathcal{F}} [\text{ driving force}] \qquad (28)$$

If, in an electrodialysis application, the electrical component of the driving force is much more significant than the diffusional component, i.e.:

$$\text{grad } (\ln \bar{C}_i \bar{\gamma}_i) << z_i \mathcal{F} \text{ grad } \phi$$

$$\text{and } J_i \doteq \pm \bar{U}_i \bar{C}_i \text{ grad } \phi,$$

then the ratio of fluxes of any two species approaches the ratio of the products of their mobilities and concentrations:

$$\frac{J_i}{J_k} \doteq \frac{\bar{U}_i \bar{C}_i}{\bar{U}_k \bar{C}_k} \; \frac{\text{grad } \phi}{\text{grad } \phi} = \frac{\bar{U}_i \bar{C}_i}{\bar{U}_k \bar{C}_k} \qquad (29)$$

We hoped this relationship might lead to a relatively simple method of examining membranes for their utility in separating similarly charged ionic species.

The specific conductance of an ion-exchange membrane may be expressed:[15]

$$\bar{K} = \frac{d}{RA} \doteq 0.001 \; \mathcal{F} \sum \bar{U} \bar{C}', \qquad (30)$$

so that the specific conductance of the membrane completely in the form of species i would be:

$$\bar{K}_i = \frac{d_i}{R_i A} \doteq 0.001 \ \mathcal{F} \ \bar{U}_i \bar{C}_i' \tag{31}$$

The mobility of the species i would then be:

$$\bar{U}_i \doteq \frac{1{,}000\ d_i}{\mathcal{F}\ \bar{C}_i \quad R_i A} \tag{32}$$

where:

d_i = thickness of swollen membrane equilibrated with species i (cm);

$\bar{C}_i'$ = milliequivalents of species i per ml of swollen membrane;

R_i = electrical resistance of membrane equilibrated with species i (ohms);

A = area of measuring probes (cm^2).

The ratio of the mobilities then of any two species will be:

$$\frac{\bar{U}_i}{\bar{U}_k} = \frac{d_i \ \bar{C}_k' \ R_k}{d_k \ \bar{C}_i' \ R_i} \tag{33}$$

If the membrane is equilibrated with only one species i, the concentration $\bar{C}_i'$ may be expressed in terms of membrane capacity;

$$\bar{C}_i' = \frac{(\text{Capacity, meq})}{(\text{Area, cm}^2) \times (d_i\text{, cm})} \tag{34}$$

Combining equations (33) and (34):

$$\frac{\bar{U}_i}{\bar{U}_k} = \left(\frac{R_k}{R_i}\right) \left(\frac{d_i}{d_k}\right)^2 \tag{35}$$

Thus, the ratio of the ionic mobilities of any two species in a particular membrane may be determined by measurements of the electrical resistance and thickness of the membrane equilibrated first with one species and then with the other.

If now, a membrane sample is equilibrated in a mixed solution of two ionic species, and the relative equilibrium concentrations are determined, an ion-exchange separation factor may be described:[13]

$$\begin{pmatrix} i \\ \alpha \\ k \end{pmatrix}_{IE} = \frac{\bar{C}_i}{\bar{C}_k} \cdot \frac{C_k}{C_i} \tag{36}$$

where:

$\bar{C}_i$, $\bar{C}_k$ = equilibrium membrane concentrations of species i and k;

C_i , C_k = equilibrium solution concentrations of species i and k;

$\begin{pmatrix} i \\ \alpha \\ k \end{pmatrix}_{IE}$ = ion-exchange separation factor.

This "ion-exchange separation factor" (equation (36)) is different from the separation factor described by equation (26) in that the former deals with static equilibrium membrane concentrations while the latter is concerned with changes in solution concentrations due to ionic transport across the membrane. Combining equations (26), (29), and (35), will yield an expression for the separation factor described by equation (37):

$$\begin{pmatrix} i \\ \alpha \\ k \end{pmatrix} = \frac{J_i/C_i}{J_k/C_k} \doteq \frac{\bar{U}_i\bar{C}_i/C_i}{\bar{U}_k\bar{C}_k/C_k} \doteq \frac{R_k}{R_i}\left(\frac{d_i}{d_k}\right)^2 \begin{pmatrix} i \\ \alpha \\ k \end{pmatrix}_{IE} \tag{37}$$

Thus, it appears that separation factors might be estimated from static equilibrium measurements. The derivation of equation (37) necessitated the following assumptions:

1. All similarly charged species are subjected to essentially the same driving force; this implies that the log-activity gradients of all similarly charged species are either similar or negligible compared to the potential gradient. If none of the separation factors differ greatly from unity, all log-activity gradients will, in fact, be similar; that the electrical potential gradient is much more significant than activity gradients in an electrodialysis system appears to be a reasonable assumption, but has not been verified experimentally.

2. Individual ionic mobilities remain constant in the presence of other species; although this is not strictly true, the differences are negligible compared to other possibilities of error.[16]

3. Convective transport is ignored; while the derivation does not specifically consider convection, the mobilities determined by resistivity measurements certainly include the effects of convective transport.

4. Static or ion exchange equilibrium concentrations do not differ from the dynamic membrane equilibrium concentrations found in an electrodialysis system; this assumption, while adequate as a first approximation, does not consider the effects of different solutions on either side of the membrane.

Until the assumptions noted are verified experimentally, separation factors calculated from static equilibrium measurements cannot be regarded quantitatively. Nevertheless, these measurements offer a qualitative description of the effects of modification of various membrane parameters upon selectivity.

1. Test Procedure

The following solutions were used in the mobility ratio determinations:

Salt	Solution 1	Solution 2
KCl	1.0N Cl^-	0.01N Cl^-
KNO_3	1.0N NO_3^-	0.01N NO_3^-
K_2SO_4	1.0N $SO_4^=$	0.01N $SO_4^=$
(pH 4.5) KH_2PO_4	1.0N $H_2PO_4^-$	0.01N $H_2PO_4^-$
(pH 9.5) K_2HPO_4	2.0N $HPO_4^=$	0.01N $HPO_4^=$
(pH 13) K_3PO_4	1.5N $PO_4^{\equiv}$	0.01N $PO_4^{\equiv}$

Membrane samples were equilibrated for 24 hours in the more concentrated solutions. This was followed by three washings in the more dilute solutions to remove any excess salt concentration from the surface or interior of the membranes. Samples were then equilibrated in the dilute solutions an additional 24 hours before resistance and thickness measurements were taken.

Samples consisted of 3/4-by 3-inch strips of four somewhat randomly selected types of anion permeable membranes. All membrane types were standard Ionics' Nepton membranes.[9] Properties of these various membrane types are given in Table 25. Two strips of each membrane type were equilibrated with each species and at least three individual resistance and thickness readings were taken with each strip.

Table 25. DESCRIPTION OF MEMBRANE PROPERTIES

Membrane type	Batch No.	Resistance, ohm-cm^2	Capacity, meq/dry gram	Water content, % weight
AR110DYG-067	347XA	12.3	2.6	32.7
AR110CZG-067	296	26.0	1.8	40.6
AR111BZL-183	184-A	15.2	1.8	45.2
AR111GWL-183	Special	23.3	1.3	53.0

For the static equilibrium measurements, mixed solutions were selected that approximated the concentration ratios found in the Clinton secondary sewage effluent used in the exhaustion runs and product-to-waste ratio studies. Total solution concentrations were approximately 0.01N.

Samples of the four selected membrane types were first placed in the Cl^- form by use of the equilibration procedure described for the mobility ratio determinations. They were then washed in distilled water and equilibrated for 7 days in the mixed solutions. These solutions were analyzed both before and after the 7-day equilibration period, and the membrane-to-solution concentration ratios calculated. Membrane samples were 3/4- by 3-inch strips. Solution volumes were 100 ml. All equilibrations were run in duplicate.

2. Test Results and Discussion

Data from the membrane parameter studies are tabulated in Appendix E.

Membrane equilibrium concentrations were determined from the changes in solution concentrations and the test strip capacities. Mobility ratios were calculated from the resistance and thickness data. By use of equation (37), separation factor estimates were then made for each of the selected membrane types. These factors are tabulated in Table 26. Considerable differences apparently exist among the inherent selectivities of the four membrane types used in this study. These membranes represent a very small sampling of the possible combinations of variables that are available and much additional effort is required to adequately investigate the influence of each parameter upon membrane selectivity. The separation factors of Table 26 are subject to the assumptions discussed in the derivation of equation (37). While inherent membrane selectivity probably can be adequately described by

Table 26. EFFECT OF MEMBRANE PARAMETERS UPON SELECTIVITY

	Membrane type			
	AR110DYG-067	AR110CZG-067	AR111BZL-183	AR111GWL-183
	Estimated separation factors			
$\alpha^{NO_3^-}_{Cl^-}$	0.5	0.8	5	8
$\alpha^{SO_4^=}_{Cl^-}$	3	3	8	7
$\alpha^{H_2PO_4^{-*}}_{Cl^-}$	28	14	7	10
$\alpha^{HPO_4^=}_{Cl^-}$	0.6	0.7	0.2	0.4
$\alpha^{PO_4^\equiv}_{Cl^-}$	2	1	26	(large)

*The equilibrium solution concentrations of $H_2PO_4^-$ were very low. The estimated separation factors therefore should be regarded as orders of magnitude only.

this equation, the selectivity obtained in operation may be quite different. A case in point concerns the phosphate - chloride separation factors for the membrane AR111BZL-183. The separation factors obtained from the membrane parameter study and from the solution parameter study are quite different. This is because the condition of negligible activity gradients assumed in the derivation of equation (37) obviously cannot be applied to the system that was used for the solution parameter study. In fact, approximately two-thirds of the combined phosphate and chloride transfer was due to diffusion because of the back diffusion of nitrate ions. The results of the two studies indicate that actual operation in a concentrating-diluting (electrodialysis) system will tend to depress the selectivities inherent in the membranes; i.e., the system will not exhibit the degree of selectivity indicated by a study of membrane parameters alone.

E. EFFECTS OF OPERATING CONDITIONS ON SELECTIVITY

Selectivity is affected by many parameters. The effects of some solution parameters, namely mole fraction and pH, upon a phosphate - chloride separation factor have been described in part C of this section; effects of membrane parameters, such as degree of cross-linkage (reflected as water content), resistance, and capacity, were illustrated in part D. Operating parameters such as the degree of approach to polarization, degree of membrane fouling or scaling, slime deposition, etc., also affect selectivity.

Analytical data from the exhaustion runs and from the product-to-waste ratio studies are tabulated in Appendix B. From these data, separation factors have been calculated according to an approximation of equation (25):

$$\begin{aligned} \alpha^{i}_{k} &= \frac{dC_i}{C_i} \cdot \frac{C_k}{dC_k} \\ &\doteq \frac{\Delta C_i}{C_i} \cdot \frac{C_k}{\Delta C_k} \\ &\doteq \frac{\%\text{ change in species i concentration}}{\%\text{ change in species k concentration}} \end{aligned} \qquad (25)$$

Rather than relate species i removal to the removal of some arbitrary reference species, individual removals have been related to total removals:

$$\alpha^{i}_{T} \doteq \frac{\%\text{ change in species i concentration}}{\%\text{ change in total solution concentration}} \qquad (6)$$

It is readily apparent that:

$$\alpha^{i}_{k} = \left(\alpha^{i}_{T}\right) \Big/ \left(\alpha^{k}_{T}\right) \qquad (38)$$

Results obtained from the separation factor approximation of equation (6) should not differ markedly from the more exact values obtained from integration of equation (25); also, more meticulous calculations might imply a greater precision of analytical data than was in fact obtained. In all calculations, percent change in total solution concentration is the percent change in total solution normality. Total solution normality is based upon analyses of important anions. Through the greatest part of the exhaustion series, sodium and potassium analyses were not available, so it was not possible to average total anion and cation equivalents. Therefore, the percent change in species i is measured in terms of normalities for all species except phosphate. Because of the variation between divalent and monovalent phosphate fractions around pH 7, percent phosphate concentration changes seemed more reasonable presented in terms of molarities. In determining total normalities, however, actual phosphate equivalents were calculated from total phosphate concentrations and pH data.

Attempts to correlate separation factor calculations with such parameters as mole fraction, pH, degree of polarization, stack resistance (as a measure of scaling or fouling), concentration ratios, time of operation, etc., have been largely unsuccessful. In view of the interdependency of the removal rates (fluxes), and therefore the separation factors of all species, it might be predicted that any very simple relationships would be unlikely. Nevertheless, certain types of data correlations have been possible which, although strictly empirical, give some insight into the effects of operating parameters upon selectivity.

The simplest type of data arrangement, a plot of percent species i removed versus percent total cut, was in most cases the most informative. For most, but not all, inorganic species studied, a fair characterization of removal rate dependency on operating conditions may be approached from this type of plot. The percent total cut is a measure of approach to polarization, or to film diffusion limiting separation factors. Because of membrane fouling, slime deposition, etc., the absolute scale may shift somewhat; i.e., a clean stack may polarize at a 50 percent cut while slime deposition may reduce this polarization point to coincide with, say, a 30 percent cut. Nevertheless, recognition of this significance of the percent total cut permits some estimate of the limits between which a particular separation factor might be expected to vary.

1. Cation Removals

Most organic materials, both suspended and dissolved, normally present in a secondary sewage effluent were assumed to be negatively charged. Organic deposits and membrane fouling were then expected on the anion membranes, rather than on the cation membranes. That this assumption was correct was indicated by organic deposits or slime growths occurring almost exclusively on the anion membranes and by the relative conditions of anion versus cation membranes at the end of the exhaustion series. No great differences in cation removal rates were anticipated between Systems I and II. It was expected that the cation membranes would function essentially as clean (instead of fouled) membranes throughout the tests, and that cation separation factor variations would primarily reflect variation of solution and operating parameters rather than membrane properties.

a. Sodium

Sodium removals for both systems are plotted against total cut in Figure 20. Most of the Sodium separation factors (determined by the slope of the lines connecting the origin to data points) appear to vary between about 0.4 for complete membrane control, to about 1.2 for film diffusion control.

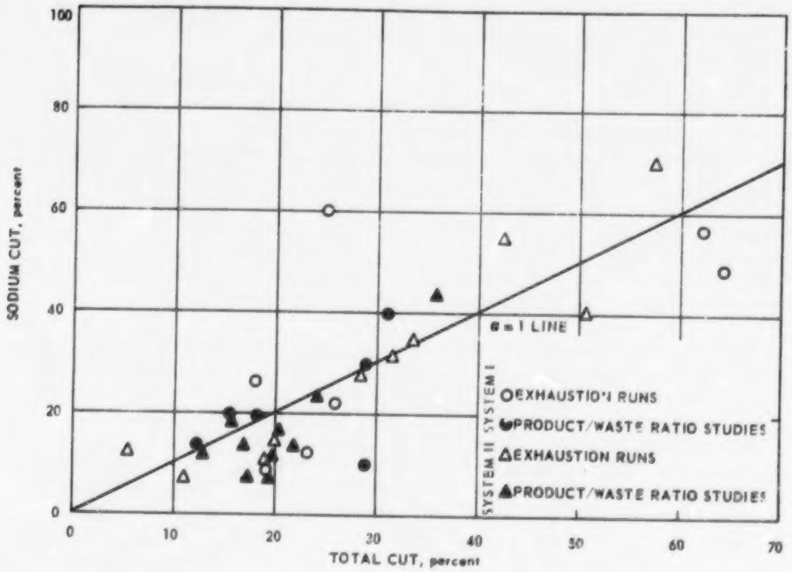


Figure 20. Sodium removals.

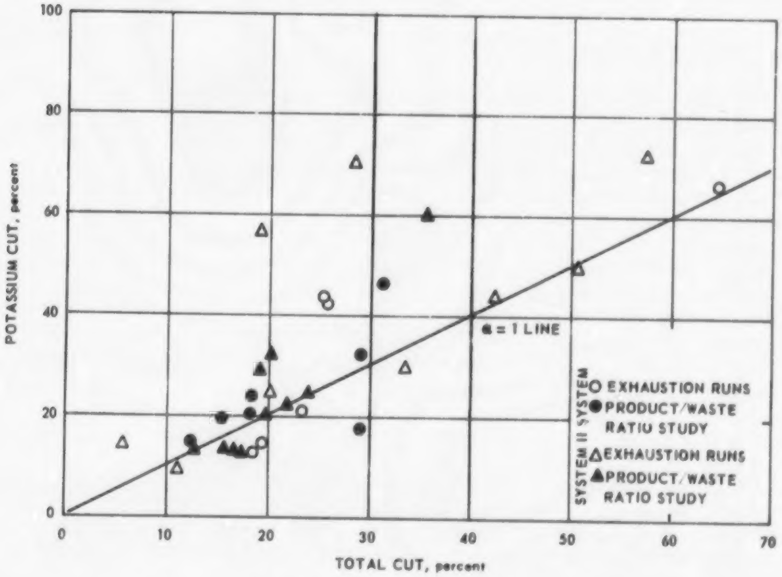


Figure 21. Potassium removals.

Between 40 and 50 percent total cut (usual design values per stack), the sodium separation factor was about 1.1.

b. Potassium

Potassium removals for both systems are plotted against total cut in Figure 21. Throughout the range of concentrations and operating conditions during the test period, the potassium separation factors did not deviate consistently from unity. From these data, a good approximation to a potassium separation factor at nearly any total cut would be 1.0.

c. Ammonium

Ammonium removals for both systems are plotted against total cut in Figure 22. Unlike removals of most other species, ammonium removals were not amenable to simplified interpretation. Between 10 and 25 percent total cut, ammonium separation factors varied from 0.2 to 2.0. This variation might be explained by the combination of fluxes of the various cationic species in this range of total removals. The tabulated analytical data in Appendix C, reveal that ammonium concentrations fluctuate over a greater range than any of the other cationic species. Since relative concentration is an important selectivity parameter, the large ammonium removal variations might possibly be explained in terms of concentration variations.

Above 30 percent total removals, ammonium separation factors begin to approach unity, and for total removals between 40 and 50 percent an ammonium separation factor of 1.0 would probably be a good approximation.

d. Calcium and Magnesium

Calcium and magnesium (total hardness) removals for both systems are plotted against total cut in Figure 23. Most calcium and magnesium separation factors appear to vary between 2.5 for complete membrane control to about 1.2 for film diffusion control. Although calcium and magnesium removals were depressed at times because of scaling in concentrate streams, the separation factor was still nearly always greater than unity.

2. Anion Removals

The type of plot used to characterize anion removals is similar to that for cation removals. In addition to showing the dependency of separation factors upon total cut (or approach to polarization), the anion removal plots also describe the effects of membrane fouling. In contrast with cation removals, a clear difference is noted in anion removals between Systems I and II.

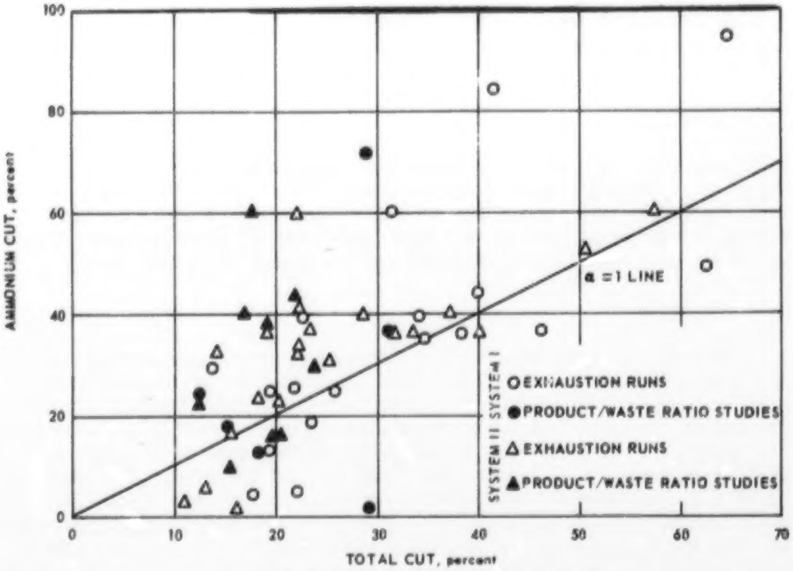


Figure 22. Ammonium removals.

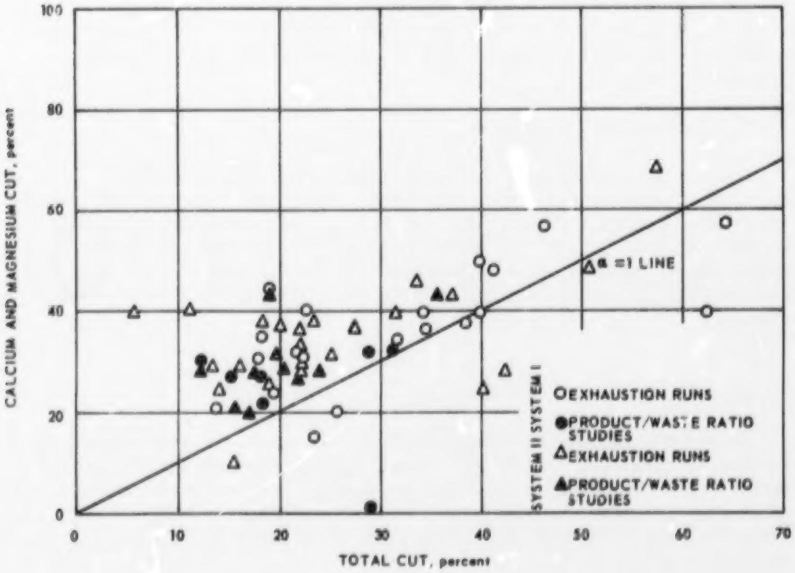


Figure 23. Calcium and magnesium removals.

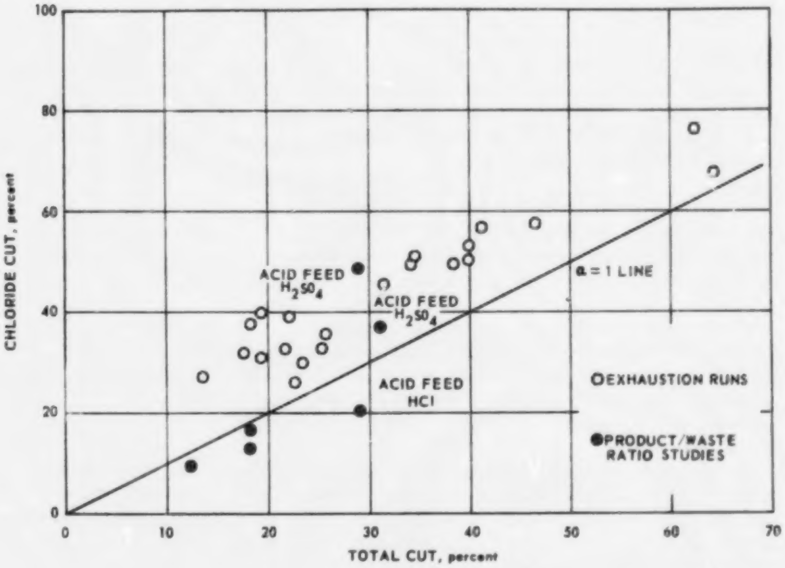


Figure 24. Chloride removals – system I.

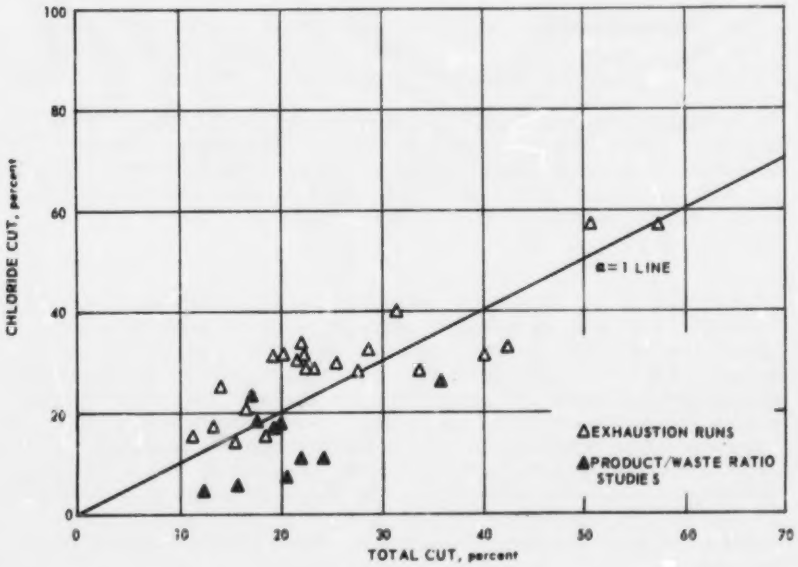


Figure 25. Chloride removals – system II.

a. Chloride

Chloride removals are plotted against total cut in Figure 24 for System I and in Figure 25 for System II. Two effects are noted immediately from these figures: fouling of anion membranes acts to increase chloride removal, and use of HCl as the concentrate-stream acid feed acts to depress chloride removal.

Figure 25, which describes relatively "clean" or non-fouled membranes, shows a chloride separation factor that normally varies from 1.3 for membrane controlled transfer to about 1.1 for film-diffusion controlled transfer (exhaustion run data only).

Data for the more fouled System I (Figure 24) show the chloride separation factor to be increased to 2.0 for membrane controlled transfer and to 1.3 or lower for film diffusion control. Over the range of conditions studied, chloride removals were consistently higher in the more fouled system.

The use of HCl as concentrate-stream acid feed in the product-to-waste ratio studies resulted in an increased chloride concentration gradient tending to resist chloride removal. Chloride separation factors from the product-to-waste ratio studies range from 0.3 to 0.8 and are consistently less than unity.

b. Bicarbonate

Bicarbonate removals are plotted against total removal in Figure 26 for System I and in Figure 27 for System II. The effect of membrane fouling appears to be a depression of bicarbonate removal.

Exhaustion run data for the relatively "clean" System II indicate a bicarbonate separation factor that varies generally between 0.6 for membrane control and about 1.1 for film diffusion control. Exhaustion run data for the more fouled System I indicate a bicarbonate separation factor varying between 0.5 for membrane control and approaching 1.0 for film diffusion control.

Acid feed to the concentrate streams increases bicarbonate removal markedly. The low concentrate-stream pH results in essentially complete conversion of bicarbonate ions to dissolved CO_2 in the concentrate stream. The resulting bicarbonate concentration gradient increases the driving force acting to remove bicarbonate species. This effect is seen in both Figures 26 and 27 for the product-to-waste ratio studies data. The bicarbonate separation factor in System II appears to approach a value around 2.0 for total cuts greater than 30 percent. The System I bicarbonate separation factor approaches 1.5 at total cuts greater than 30 percent. This difference between systems probably again reflects the effect of anion membrane fouling that tends to depress bicarbonate removal.

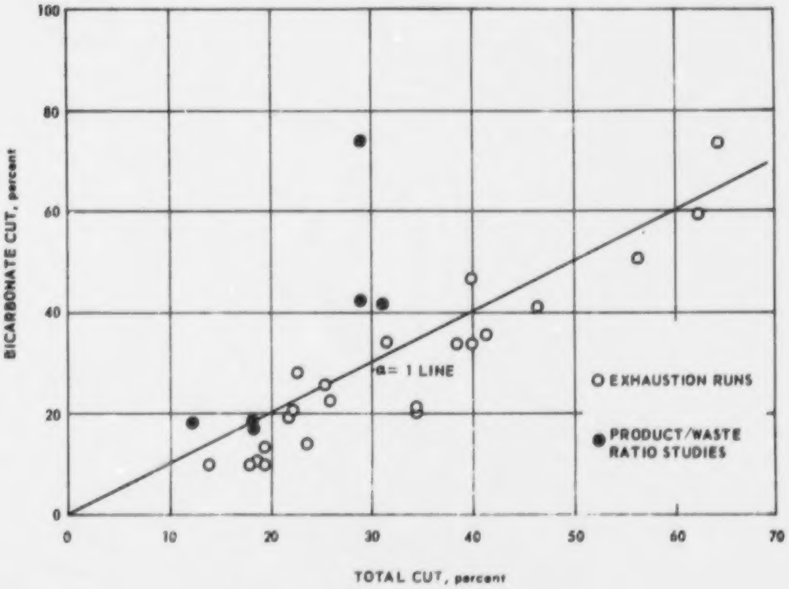


Figure 26. Bicarbonate removals – system I.

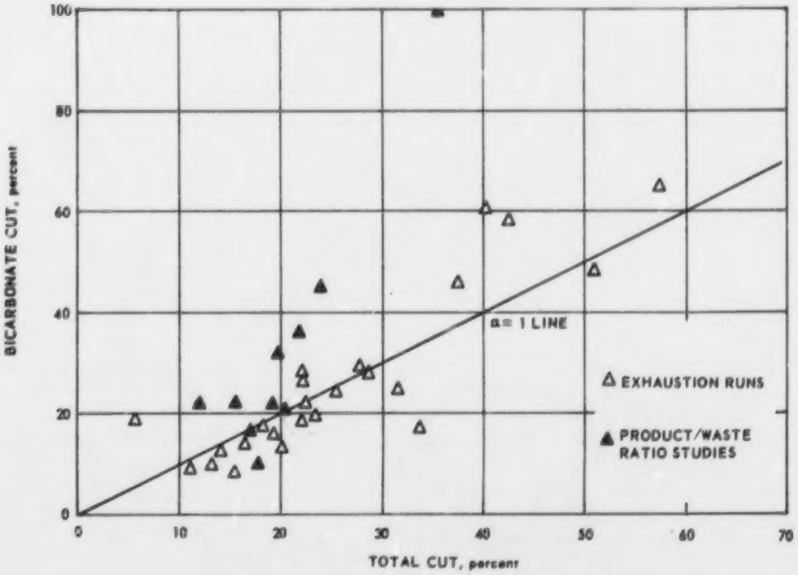


Figure 27. Bicarbonate removals – system II.

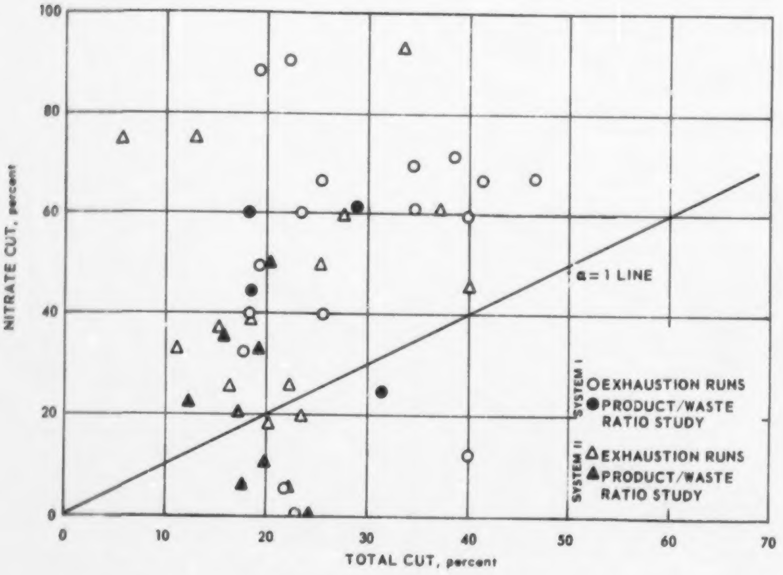


Figure 28. Nitrate removals.

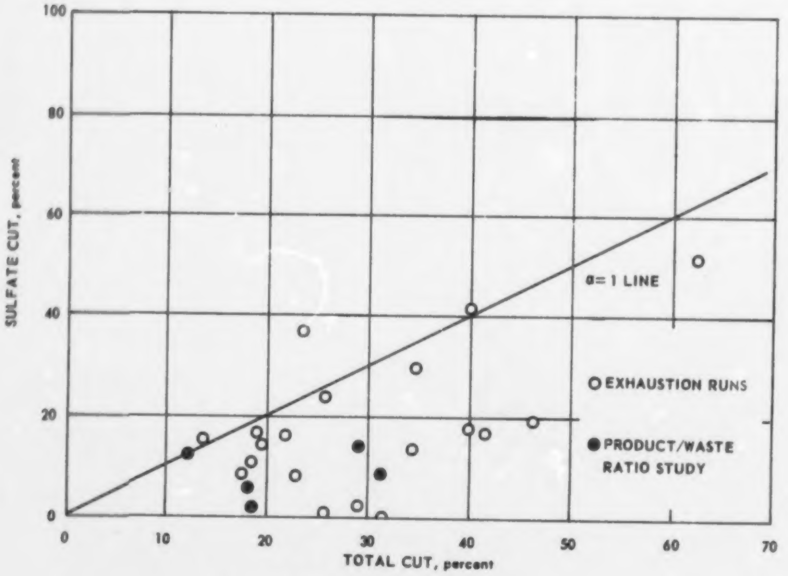


Figure 29. Sulfate removals – system I.

c. Nitrate

Nitrate removals are plotted against total cut in Figure 28 for both System I and System II. Nitrate removals are difficult to characterize in the area of membrane controlled transfer. Data in Appendix C show that nitrate concentrations vary tremendously. This conceivably could account for the nitrate removal variations experienced. Beyond a total cut of about 20 percent, nitrate removal data became more consistent and appeared (for System I at least) to approach a separation factor value of around 1.5 for film diffusion controlled transfer. For total cuts greater than about 20 percent, nitrate separation factors were with few exceptions greater than unity for both systems. For both systems, the maximum consistent value appeared to approach 2.5, which was attained at about a 20 percent total cut.

d. Sulfate

Sulfate removals are plotted against total cut in Figure 29 for System I and in Figure 30 for System II. Sulfate removals appear erratic in both systems. While chloride and bicarbonate removals can be treated quantitatively, sulfate removals cannot. Qualitatively, the data from the "clean" System II indicate a sulfate separation factor approaching a value slightly greater than unity for film diffusion control. The data from the more fouled System I do not so indicate. In fact, nearly all

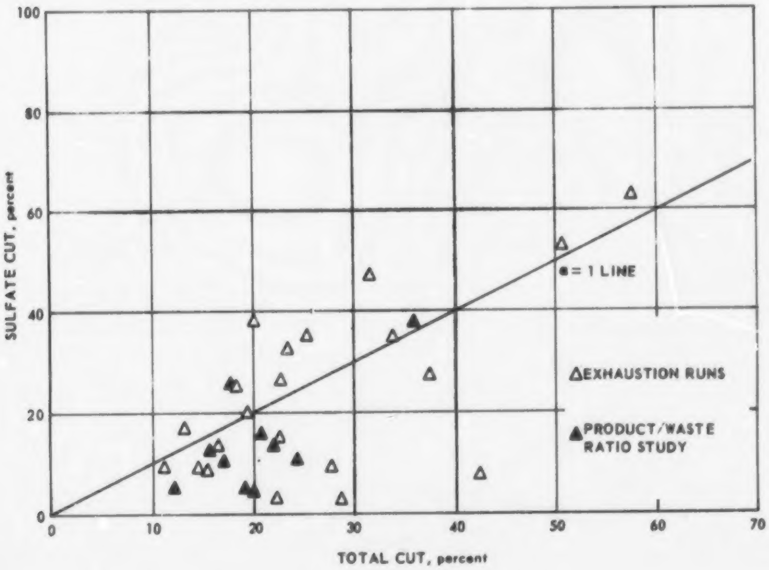


Figure 30. Sulfate removals. – system II.

data points at any total-cut value are below the =1 line. This indicates that sulfate removal is depressed by membrane fouling, a fact which is borne out by electrodialysis field experience with sulfate-containing well waters.

e. Phosphate

Phosphate removals appear to be the most affected by membrane fouling. The most revealing display of phosphate-removal data is the plots of phosphate separation factors against time of operation (Figures 31 and 32). Phosphate separation factors as high as 2.0 were realized with clean stacks. With operating time (and increased fouling) the phosphate separation factors decreased gradually over an order of magnitude. This decrease, moreover, was much more rapid with System I, which was fouled more rapidly.

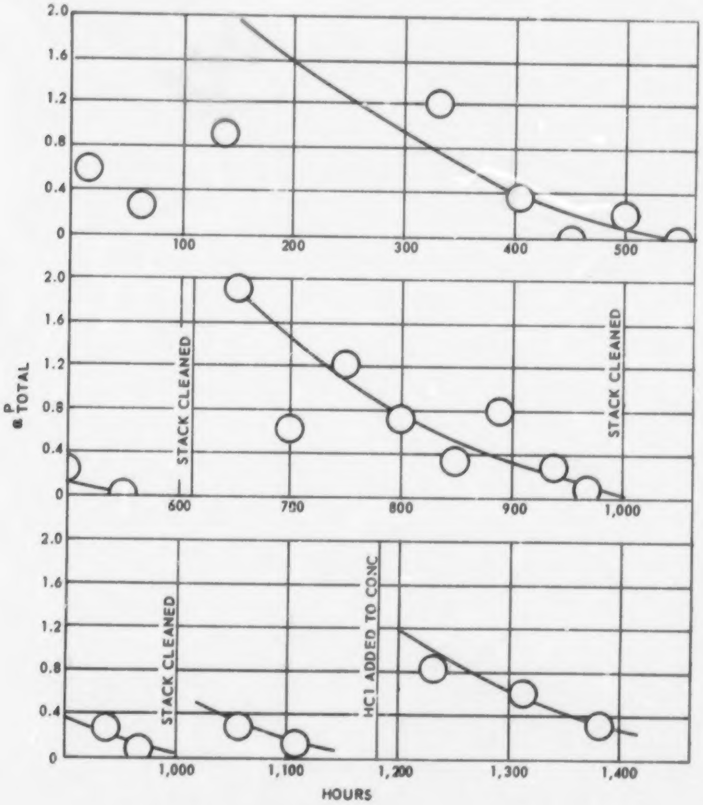


Figure 31. Phosphate removal – exhaustion runs, system I (α^P_{Total} vs. time of operation).

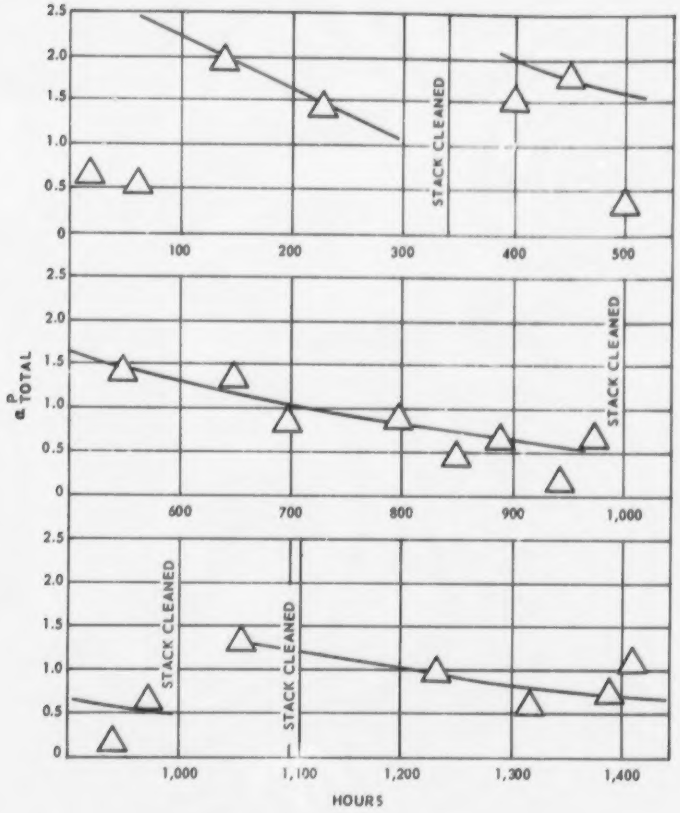


Figure 32. Phosphate removal – exhaustion runs, system II (α^{P}_{Total} vs. time of operation).

SECTION IV
PRELIMINARY DESIGN AND COST ESTIMATE OF 10-MGD ELECTRODIALYSIS PLANT

A. DEVELOPMENT OF DESIGN EQUATIONS

The design of electrodialysis equipment has been described in several publications.[12,15,17,18] These design procedures are based upon the application of Faraday's and Ohm's Laws. The removal of inorganics is defined by Faraday's Law:

$$\frac{\text{gram equivalents transferred}}{\text{sec}} = \frac{iA_p\ (CE)\ n}{1{,}000\ \mathcal{F}} \tag{39}$$

where:

i = current density, ma/cm^2;

A_p = effective transfer area, cm^2/cell-pair;

CE = current efficiency;

n = number of cell-pairs;

$\mathcal{F}$ = Faraday's constant, 96,500 coulombs/equivalent.

Energy consumption is described by Ohm's Law:

$$\text{Energy, } \frac{\text{kw-hr}}{1{,}000\text{ gal}} = \frac{0.00105\ i^2\ A_p^{\ 2}\ R_s}{10^6\ Q_T} \tag{40}$$

where:

R_s = electrical resistance of membrane stack, ohms;

Q_T = diluting stream flow rate, liters/sec.

Electrical resistance of membrane systems usually is described in units of ohm-cm^2/cell-pair. Correlation of electrical resistance with an averaged solution normality is accomplished with the relationships:[12]

$$\frac{1}{C_a} = \frac{1}{2}\left(\frac{1}{C_d} + \frac{k}{C_c}\right) \tag{41}$$

$$k = t_c/t_d \qquad (42)$$

$$R_p = \frac{K_1}{C_a} + K_2 - K_3 C_a \qquad (43)$$

where:

C_a = averaged normality of diluting and concentration streams, C_d and C_c;

t_c = thickness of concentrating stream;

t_d = thickness of diluting stream;

R_p = cell-pair resistance, ohms-cm^2/cell-pair.

K_1, K_2 and K_3 are constants depending upon various membrane, solution and temperature parameters. K_1/C_a is a solution resistance term; K_2 is a membrane resistance term; and K_3C_a accounts for non-linearity of solution and membrane resistances with concentration at very high electrolyte concentrations. In a waste-water reclamation application, solutions will be relatively dilute, so the K_3C_a term will be insignificant. Cell-pair resistances will then be:

$$R_p \doteq \frac{K_1}{C_a} + K_2 \qquad (44)$$

Following the procedure of Mason and Kirkham[12] for parallel-flowing concentrating and diluting streams, the concentrations at any point L along a flow path would be

$$C_d = C_{di} \; (1 - f) \qquad (45)$$

and:

$$C_c = C_{ci} + \frac{F_d}{F_c} \; C_{di} \; f \qquad (46)$$

where:

f = fractional cut to point L;

C_{di}, C_{ci} = influent dilute and concentrate stream concentrations, respectively;

F_d,F_c = dilute and concentrate stream flow rates, respectively, liters/sec.

With the definitions:

$$g = C_{di}/C_{ci} \tag{47}$$

and

$$\beta = F_c/F_d \tag{48}$$

the averaged normality may be rewritten:

$$C_a = 2C_{di} \left\{ \frac{(1 - f)\ (\beta + fg)}{\beta(1 + gk)\ +\ fg\ (1 - \beta k)} \right\} \tag{49}$$

and the cell-pair resistance expression may be rewritten:

$$R_p = \frac{K_1}{2C_{di}} \left\{ \frac{\beta\ (1 + gk)\ +\ fg\ \ (1 - \beta k)}{(1 - f)\ \ (\beta + fg)} \right\} + K_2 \tag{50}$$

The fractional cut per unit flow path length at point L is:

$$\frac{df}{dL} = \frac{W\ \ (CE)\ i}{1{,}000\ F_d\ C_{di}\ \mathcal{F}} \tag{51}$$

where:

W = flow path width, cm.

The potential drop at point L is:

$$V_p = \frac{i\ R_p}{1{,}000} \tag{52}$$

where:

V_p = voltage drop per cell pair.

Combining equations (51) and (52):

$$\frac{V_p W\ \ (CE)}{F_d\ C_{di}\ \mathcal{F}}\ dL = R_p\ df \tag{53}$$

Substituting the resistance equation (50) and assuming constant current efficiency, equation (53) may be integrated over $L = 0$ to $L = L_o$ and $f = 0$ to $f = f$. This integration yields the design equation:

$$\frac{V_p A_p (CE)}{F_d \quad \mathcal{F}} \quad \frac{K_1}{2}\left[\frac{1+gk}{1+\alpha} \ln\left(\frac{1+\alpha f}{1-f}\right) - \left(\frac{\alpha - gk}{1+\alpha}\right) \ln (1-f)(1+\alpha f)^{1/\alpha}\right] + K_2 f C_{di} \tag{54}$$

where:

$$\alpha = \frac{g}{\beta} = \frac{F_d C_{di}}{F_c C_{ci}} \tag{55}$$

For the case of equal spacer thicknesses, $k = 1$, and equal flow rates, $\beta = 1$ and $\mathcal{F} = g$, equation (54) will reduce to:

$$\frac{V_p A_p (CE)}{F_d \quad \mathcal{F}} = \frac{K_1}{2} \left(\ln \frac{(1+gf)}{(1-f)}\right) + K_2 f C_{di} \tag{56}$$

As discussed in Section II of this report, the ratio i/C_d is a measure of approach to polarization. Thus there will be a limiting value for this ratio, beyond which equation (56), based upon constant current efficiency, will fail to hold true. The i/C_d ratio at the outlet of the diluting stream is:

$$\frac{i}{C_d}{}_o = \frac{i_o}{C_{do}} = \frac{1{,}000\ V_p}{(R_p)_o C_{do}} = \frac{1{,}000\ V_p}{(R_p)_o C_{di} \ (1-f_o)} \tag{57}$$

Substituting this relationship in equation (56) yields:

$$\frac{A_p (CE) \ (i/C_d)_o}{F_d \quad \mathcal{F}} = \frac{1{,}000}{(1-f)} \left[\frac{\frac{K_1}{2 C_{di}} \ln \frac{(1+gf)}{(1-f)} + K_2 f}{\frac{K_1}{2 C_{di} (1-f)} \frac{(1+g)}{(1+gf)} + K_2}\right] \tag{58}$$

Rearranging:

$$A_p = \frac{1{,}000\ F_d \mathcal{F} [2K_2 C_{di} f + K_1 Z]}{(CE)\ (i/C_d)_o \left[2K_2 C_{di} (1-f) + K_1 Y\right]} \tag{59}$$

where:

$$Y = 1 + \frac{(1-f)g}{gf+1} \tag{60}$$

$$Z = \ln \frac{(1+gf)}{(1-f)} \tag{61}$$

Similarly, equation (57) may be rearranged:

$$V_p = \frac{(i/C_d)_o}{1{,}000} \left[\frac{K_1}{2} Y + K_2 (1-f) C_{di} \right] \qquad (62)$$

Equations (59) and (62) are the basic design equations that will be used to determine the required size and energy consumption of an electrodialysis installation.

B. DESIGN CRITERIA

In the preliminary design calculations for a waste-water reclamation system, feed composition was assumed to approximate that of the averaged secondary-sewage-treatment effluents as compiled by the United States Public Health Service in their Advanced Waste-Treatment Research Program. The design composition (Table 27) is similar to that

Table 27. DESIGN COMPOSITION

Dilute Stream Influent

Constituent	Concentration, mg/l
Sodium, Na^+	130
Potassium, K^+	15
Ammonium, NH_4^+	20
Calcium, Ca^{+2}	60
Magnesium, Mg^{+2}	25
Chloride, Cl^-	115
Nitrate, NO_3^-	10
Bicarbonate, HCO_3^-	300
Sulfate, SO_4^{-2}	100
Silica, SiO_3^{-2}	50
Phosphate, PO_4^{-3}	25
Total Dissolved Solids	850

pH = 7.4

Solution normality ≐ 10 meq/liter $C_{di} = 0.010N$

used in the exhaustion runs and in the product-to-waste ratio studies except that the sodium, chloride, bicarbonate, and sulfate concentrations are approximately tripled.

Table 28. MAXIMUM ACCEPTABLE CONCENTRATIONS[a]

Constituent	Concentration, mg/l
Sodium, Na^+	110
Potassium, K^+	50
Ammonium, NH_4^+	1
Calcium, Ca^{+2}	50
Magnesium, Mg^{+2}	30
Chloride, Cl^-	250
Nitrate, NO_3^-	45
Bicarbonate, HCO_3^-	150
Carbonate, CO_3^{-2}	20
Sulfate, SO_4^{-2}	250
Silica, SiO_3^{-2}	50
Phosphate, PO_4^{-3}	200
Hardness, $CaCO_3$	125
Alkalinity, $CaCO_3$	250
Total Dissolved Solids	500

[a] These values were suggested by the Advanced Waste-Treatment Research Program for purposes of this report and are not official standards.

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For this report, a set of unofficial "maximum acceptable concentrations" of inorganic contaminants in the product water was suggested by the Advanced Waste-Treatment Research Program of the Public Health Service (Table 28). A preliminary design was based upon a reduction of total dissolved solids (TDS) from 850 to 500 ppm. Separation factors determined in the test operations indicated that "acceptable" ammonium concentrations would not be realized with this reduction. Therefore, ammonium concentrations of less than 1 ppm presumably will be attained through some auxiliary process, e.g., increased nitrification in the secondary treatment stage and/or adsorption of molecular ammonia in the activated carbon stage. A more refined analysis might take into account the individual increments added per pass through the municipality, the effect of concentration upon separation factors, the effect of additions to the system from a source stream, etc. At this point, however, without more basis selectivity information and some information about the fluctuations of flow and concentration with time, the introduction of these additional considerations would be both confusing and unjustified. The design value of the fractional cut f, then, will be

$$f = \frac{850-500}{850} = 0.41.$$

Dilute- and concentrate-stream spacer geometry is assumed to be identical (the concentrate stream operating on a feed-and-bleed basis). Thus:

$$k = \beta = 1 \quad \text{and} \quad \alpha = g,$$

so equations (59) through (62) may be used for the system design.

A g-value of 0.15 is used. This corresponds to a waste-to-product concentration ratio of:

$$\frac{C_{co}}{C_{do}} = \frac{6{,}010}{500} = 12/1$$

This is a conservative design value based upon field experience and the product-to-waste ratio studies described in Section II of this report. Since $F_d = F_c$, the product-to-waste flow ratio in this case would be:

$$\frac{F_d}{F_w} = \frac{5{,}660 - 500}{850 - 500} \doteq 15/1$$

or about 7 percent waste stream.

Based upon the exhaustion runs, product-to-waste ratio studies, and electrodialysis field experience, current efficiency is estimated at 90 percent.

Stack velocities will be about 50 centimeters per second. This will be achieved, if necessary, by alteration of the number of parallel flow paths in the spacer design.

Figure 17 of this report indicates that the limiting ($^{i}/C$) value at 50 centimeters per second will be about 1000, even with moderate fouling or slime deposits. The design $(^{i}/C)_o$, reduced by 25 percent to include a factor of safety, will be 750.

Water temperature will be assumed to be 50°F (10°C).

With adequate maintenance and competent operating personnel, anticipated membrane life will be 5 years, i.e., 20 percent replacement every year.

Capital investment is assumed to be amortized at 5 percent over 25 years.

C. DESIGN CALCULATIONS

1. Stack Requirements

$$\frac{1}{C_a} = \frac{1}{2}\left(\frac{1}{C_d} + \frac{1}{C_c}\right) ; \quad (k = 1)$$

Since $C_c >> C_d$, then:

$$\frac{1}{C_a} \doteq \frac{1}{2C_d} \quad \text{or} \quad C_a = 2C_d$$

From Appendix D:

$$@\ 25°\text{C:} \quad \rho \quad (\text{ohm-cm}) = \frac{9.5}{C_d}$$

or, using a flow path thickness of 0.1 cm,

$$R\ (\text{ohm-cm}^2) = \frac{9.5}{C_d} \quad (0.1\ \text{cm}) = \frac{0.95}{C_d}$$

Assuming T = 10°C:

$$R\ (\text{ohm-cm}^2) = \frac{0.95}{C_d}\left[1 + 0.02\ (25-10)\right] = \frac{1.25}{C_d}$$

or

$$R\ (\text{ohm-cm}^2) = \frac{1.23}{C_a} \times 2 = \frac{2.46}{C_a}$$

This value is increased by 30 percent to allow for the boundary film contribution to resistance, so:

$$R = \frac{2.46}{C_a}\ (1.30) = \frac{3.2}{C_a}$$

Thus, 3.2 corresponds to K_1 in equations (59) and (62).

The K_2 value is the sum of the anion and cation membrane resistances. Tables 14 and 15 in Section II show this to be 77.4 ohm-cm^2 for System I or 26.0 for System II. Although adequate pretreatment of the waste stream would prevent the membrane resistances reaching those of the fouled System I, to be conservative this higher value is used in the design.

Thus, K_2 = 77.4 ohm-cm^2
From equation (60):

$$Y = 1 + \frac{(1-f)g}{gf + 1} = 1 + \frac{(0.59)\ (0.15)}{(0.15)\ (0.41) + 1} = 1.08$$

From equation (61):

$$Z = \ln\left(\frac{1 + gf}{1 - f}\right) = \ln\ 1.80 = 0.588$$

Using design equation (59), and realizing that total transfer area equals nA_p, where n equals the number of cell-pairs:

$$nA_p = \frac{10^3\ nF_d\ \mathcal{F}}{(CE)\ (i/C_d)_o}\left[\frac{2K_2\ C_{di}f + K_1Z}{2K_2\ C_{di}\ (1-f) + K_1Y}\right]$$

From Table 27, $C_{di} = 0.01N$; so:

$$nA_p = \frac{10^3(4.37 \times 10^2)(96{,}500)}{(0.9)(750)} \left[\frac{2(77.4)(0.01)(0.41) + (3.2)(0.588)}{2(77.4)(0.01)(0.59) + (3.2)(1.08)}\right]$$

$$= 624 \times 10^5 \left[\frac{2.515}{4.364}\right] = 35.8 \times 10^6 \text{ cm}^2$$

or:

$$\frac{35.8 \times 10^6}{929} = 38{,}600 \text{ ft}^2$$

This is the total required effective cell-pair area.

The Ionics' Mark III electrodialysis stack has an effective cell-pair area of 3,100 square centimeters so, with this type of equipment:

$$\frac{35.8 \times 10^6 \text{cm}^2}{3{,}100 \text{ cm}^2\text{/cell-pair}} = 11{,}530 \text{ cell-pairs required.}$$

The standard Mark III has 275 cell-pairs per stack, so:

$$\frac{11{,}530 \text{ cell-pairs}}{275 \text{ cell-pairs/stack}} = 42$$

or: 42 Mark III stacks required.

The plant will be designed with 44 such stacks to allow for safety and periodic removal of stacks from operation for routine maintenance and cleaning.

A check on stack velocities shows:

$$\frac{4.37 \times 10^2 \text{ liter/sec}}{42 \times 275 \text{ cell-pairs}} \times 1{,}000 = 37.9 \frac{\text{ml/sec}}{\text{cell-pair}}$$

A standard Mark III spacer, 1 mm thick, with eight parallel fl
paths will approximate the desired 50 centimeters per second ve
at this flow rate.

2. Power Requirements

The stack voltage requirements are determined by means of design equation (62):

$$V_p = \frac{(i/C_d)_o}{1,000} \left[\frac{K_1}{2} Y + K_2 (1-f) \; C_{di}\right]$$

$$= \frac{750}{1,000} \left[\frac{3.2}{2} (1.08 + (77.4) (0.59) (0.01)\right]$$

$$= 0.75 \times 2.184 = 1.64 \text{ volts/cell-pair}$$

Stack voltage = (1.64) (275) = 451 volts/stack.

DC current is calculated from the current efficiency equation:

$$I = \frac{Q_d \; \mathcal{F} \; f \; C_{di}}{n \, CE}$$

$$= \frac{(0.0379) \; (96,500) \; (0.41) \; (0.01)}{(0.90)}$$

$$= 16.7 \text{ amperes/stack}$$

The DC power requirement is:

$$P = VI$$

$$= (451 \text{ volts}) \; (16.7 \frac{\text{amperes}}{\text{stack}}) \; (42 \text{ stacks})$$

$$= 316 \text{ kv-a}$$

Electrode voltage requirements are not included in this power requirement. If we assume a total electrode voltage per stack of 3 volts for the electrolysis of water and the overvoltages of platinum and stainless steel electrodes, the electrode power requirement is:

$$(3\text{ volts})\ (16.7\ \frac{\text{amperes}}{\text{stack}})\ (42\text{ stacks}) = 2100\text{ watts.}$$

Total DC power requirement then is:

$$P_T = 316 + 2.1 = 318.1 \doteq 320\text{ kv-a}$$

A 500-kv-a rectifier would be used.

3. Acid Feed Requirements.

To prevent carbonate scaling of the concentrate compartments, sufficient acid is added to the concentrate stream to force the reaction:

$$HCO_3^- + H^+ \longrightarrow H_2CO_3$$

Since approximately half the bicarbonate content of the dilute streams is transferred to the concentrating compartments, an equivalent amount of acid is added to these compartments. Other techniques are sometimes used to prevent scaling, 6,19 but for preliminary design and cost estimates, calculations for acid feed will suffice. The acid requirement is:

$$(2.5\ \frac{\text{meq}}{\text{liter}}\ HCO_3^-\ \text{removed})\ (3.785\ \frac{\text{liter}}{\text{gal}}) = 9.5\ \frac{\text{meq}}{\text{gal}}$$

$$\text{or}\quad 9.5\ \frac{\text{equiv}}{1{,}000\text{ gal}}$$

Because of hydroxide production, an additional amount of acid is required for the cathode stream; however, use of the acidic effluent from the anode compartments as cathode compartment influents will greatly reduce the cathode acid feed.

4. Calcium Sulfate Solubility

Since the product-to-waste flow ratio will be about 15 to 1, the sulfate and calcium concentrations in the concentrate stream will be 15 times their concentration reductions in the product stream. In addition, the sulfate concentration in the concentrate stream will be increased by the acid treatment (if H_2SO_4 is used).

The following separation factors are assumed from the results of the product-to-waste ratio studies:

$$\alpha \frac{Ca^{+2}}{Total} \doteq 1.1$$

$$\alpha \frac{SO_4^{=}}{Total} \doteq 0.5$$

With these separation factors and the influent concentrations specified in Table 27, the following waste stream concentrations are calculated:

$$Ca^{+2} : (\frac{15}{1}) \ (60 \frac{mg}{liter}) \ \frac{(0.41)\ (1.1)}{(40 \frac{gm}{mole})\ (1000 \frac{mg}{gm}} = 0.01M$$

$$SO_4^{=} : (\frac{15}{1}) \ \frac{(100 \frac{mg}{liter})\ (0.41)\ (0.5)}{(96 \frac{gm}{mole})\ \ (1000 \frac{mg}{gm})} = 0.003M$$

In addition, if H_2SO_4 is used for acid feed,

$$\frac{15}{1} \quad \frac{(2.5 \frac{meq}{liter})\ \ (1/2 \frac{mole}{equiv})}{(1{,}000 \frac{meq}{equiv})} = 0.019M \ SO_4^{=}$$

will be added to the waste stream as acid.

Total concentrations of Ca^{+2} and SO_4^{-2} then are:

Ca^{+2} : 0.01M

$SO_4^{=}$: 0.003 + 0.019 = 0.022M

Calcium sulfate solubility product = 2.4×10^{-4}

actual $(Ca^{+2})\ (SO_4^{-2}) = (0.01)\ (0.022) = 2.2 \times 10^{-4}$

Thus, Ca SO_4 precipitation does not appear to be a problem for the proposed system.

D. COST ESTIMATE

1. Capital Investment

Stack Cost:

Membranes	$\left(\frac{3100}{929}\ \frac{\text{ft}^2}{\text{cell-pair}}\right)$ (275 x 44 cell-pairs) (\$5.80/ft^2)	\$234,000
Spacers	$\left(\frac{3100}{929}\right)$ (275 x 44) (\$1.94/ft^2)	78,500
End plates	$\left(\frac{\text{2 end plates}}{\text{stack}}\right)$ (44 stacks) (\$60.00/end plate)	5,300
Electrodes	(44) (\$815.00/pair)	35,900
Stack hardware (tie rods, etc.)		4,100
	Cost of stack components	\$357,800
	10 percent installation cost	35,800
	TOTAL STACK COSTS	\$393,600

Rectifier Cost:

(500 kv-a) (\$67.50/kv-a)		\$ 33,800
	10 percent installation cost	3,400
	TOTAL RECTIFIER COST	\$ 37,200

Auxiliary Equipment (installed): [a]

Pumps	$ 88,000
Valves, pipes, fittings	82,300
Conduit	4,900
Control panels	16,800
Control valves	16,500
Acid tanks	13,000
Cost of auxiliary equipment	$221,500
10 percent miscellaneous	22,200
TOTAL AUXILIARY EQUIPMENT COST	$243,700

Spare Parts: [b]

Stacks	$ 35,800
Rectifier	3,700
Auxiliaries	17,400
TOTAL SPARE PARTS COST	$ 56,900

Treatment Building: [c]

(6,800 ft^2) ($13.00/ft^2)

BUILDING COST	$ 88,400

Substation:

(500 kv-a) ($35.00/kv-a)

SUBSTATION COST	$ 17,500

[a] Installation cost is 40 percent

[b] Spare equipment is based upon 10 percent of uninstalled equipment cost.

[c] Building is based upon floor space requirements for Ionics' field installations.

Total Capital Investment:

Stacks	$393,600
Rectifier	37,200
Auxiliary equipment	243,700
Spare parts	56,900
Treatment building	88,400
Substation	17,500
Total	$837,300
Insurance, etc., (approximately 5 percent)	41,900
TOTAL CAPITAL INVESTMENT	$879,200

2. Amortized Capital Cost

Total capital investment is amortized at 5 percent over 25 years. Investment retirement cost then equals 7.1 percent per year.

So:

$$\text{Amortized capital cost} = \frac{(879{,}200)(0.071/\text{year})}{(365\ \text{days/year})(10^4\ 1{,}000\ \text{gal/day})} =$$

1.71 cent/1,000 gal

3. Operating Costs

DC Energy:

Use 90 percent rectifier efficiency:

$$\text{DC Energy} = \frac{(320\ \text{KW})(24\ \text{hrs/day})}{(0.90)(10^4 \times 10^3\ \text{gal/day})}$$

$$= 0.85\ \text{kw-hr/1,000 gal}$$

Use $0.01/kw-hr:

DC Energy Cost = 0.85 cents/1,000 gal

Pumping Energy:

At a 50-centimeter-per-second flow velocity, the pressure drop across a Mark III stack amounts to approximately 100 feet of water. If 80 percent efficiency is assumed for both pumps and motors:

$$\text{Pumping Energy} = \frac{\left(\frac{8337\ \text{lb}}{1{,}000\ \text{gal}}\right)\ (100\ \text{ft})\ \left(3.766 \times 10^{-7}\ \frac{\text{kw-hr}}{\text{ft-lb}}\right)}{(0.80)\ (0.80)}$$

$$= 0.491\ \text{kw-hr}/1{,}000\ \text{gal}$$

Since two streams (concentrate and dilute) are used, and energy cost is 1 cent/kw-hr:

$$\text{Pumping energy cost} \doteq 1\ \text{cent}/1{,}000\ \text{gal}$$

Acid Requirement: 9.5 equiv/1,000 gal

H_2SO_4 (66°Be´) = 8.64 eq/lb at 1 cent/lb

$$\frac{(9.5\ \text{equiv}/1{,}000\ \text{gal})(1\ \text{cent/lb})}{(8.64\ \text{equiv/lb})} = 1.1\ \text{cent}/1{,}000\ \text{gal}$$

Other Power:

Lights, etc. = 0.3 cent/1,000 gal

4. Maintenance Costs

Membrane and Spacer Replacement: [a]

Membranes	($222,800/set) (4)	=	$ 891,200
Spacers	($ 74,600/set) (4)		298,400
	Total	=	$1,189,600

$$\frac{(\$1{,}189{,}600)}{(25\ \text{yrs})(365\ \text{days/yr})(10^4\ 1{,}000\ \text{gal/day}} = 1.30\ \text{cent}/1{,}000\ \text{gal}$$

Anode Maintenance:

$$\frac{(\$200/\text{stack year})\ (42\ \text{stacks})}{(10^4\ 1{,}000\ \text{gal/day})(365\ \text{days/year})} \quad = \quad 0.23\ \text{cent}/1{,}000\ \text{gal}$$

[a] 4 sets over 25 years

Other Maintenance:

One-and-a-half percent of capital investment per year is assumed.

$$\frac{(879{,}200)\ (0.015/\text{year})}{(365\ \text{days/year})\ (10^4\ 1{,}000\ \text{gal/day})} \quad = \quad 0.36\ \text{cent}/1{,}000\ \text{gal}$$

Labor:

Electrodialysis field experience in Ionics, Incorporated, demineralization plants indicates a labor requirement of about 2 man-hours per day plus 100 man-hours per stack-year.

$$\text{Labor cost} = \frac{\left[(2)\ (365)\ +\ (100)\ (42)\right]}{(365)\ (10^4)}\ (\$4.00/\text{hr}) = 0.54\ \text{cent}/1{,}000\ \text{gal}$$

5. Cost Estimate Summary

Influent stream	Secondary sewage effluent (Pretreated for organic and suspended solids removal)
Plant capacity	10 MGD
Demineralization	850 to 500 ppm (41% cut)
Plant configuration	44 Mark III stacks in parallel
Capital investment	$879,200
Amortized capital cost	1.71 cent/1,000 gal

Operating cost		
DC energy	0.85	
Pumping	1.00	
Chemicals (H_2SO_4)	1.10	
Other power	0.30	
Total operating cost		3.25
Maintenance:		
Membrane and spacer replacement	1.30	
Anode maintenance	0.23	
Other maintenance	0.36	
Labor	0.54	
Total maintenance cost		2.43
TOTAL ESTIMATED COST		7.39 cent/1,000 gal

SECTION V
SUMMARY AND CONCLUSIONS

A. EXHAUSTION RUNS

Two exhaustion series and a product-to-waste concentration ratio study were conducted to determine the effects of a municipal secondary sewage effluent upon semi-permeable ion-exchange membranes in an electrodialysis system. Secondary sewage effluent was obtained twice weekly from the City of Clinton, Massachusetts. Nepton type AR111BZL-183 anion permeable membranes and type CR61AZL-183 cation permeable membranes, manufactured by Ionics, Inc., of Cambridge, Massachusetts, were used in these tests. Two electrodialysis systems were operated in parallel: the System I electrodialysis stack was preceded only by a cellulose cartridge filter for the removal of grit, sand, and similar microscopic particles from the secondary effluent; System II was operated in the same manner, but, in addition was also preceded by an activated carbon adsorption column for the removal of dissolved organic substances.

The first exhaustion series consisted of about 500 hours of operation for System I and about 360 hours for System II. Electrical resistance was consistently greater in System I during the latter part of this series. Examination of the electrodialysis stacks at the end of series 1 revealed large slime deposits in the diluting compartments of System I and only slight deposits in those of System II. Calcium scaling was found in concentrating compartments of both systems. Membrane resistivities had increased in both systems; some capacity deterioration was noted in anion membrane samples from System I.

Calcium scaling was due to polarized operation and the lack of pH control in the concentrate streams.

In System I, increases in membrane resistivity were not as significant as the increase in total stack resistance. This increase was attributed primarily to slime or organic deposits on the anion membranes. The increases in anion membrane resistivities were attributed to sorption of organic anions. Increases in cation membrane resistivities were attributed to iron fouling.

The second exhaustion series, in which a new set of membranes was used, consisted of about 1500 hours of operation over a 5-month period of continuous exposure to secondary effluent. During this series, we attempted to operate both stacks at about 75 percent of the limiting current density. Flow path velocities were held constant at about 6.5 cm/sec.

The second exhaustion series was followed by a 2-month product-to-waste concentration ratio study, during which concentration ratios, flow path velocities, current densities, and solution pH's were varied over wide ranges to obtain electrodialysis system performance data under a variety of operating conditions.

Electrodialysis stacks in both systems were disassembled and membranes washed at about 500-hour (operating time) intervals. At the end of the second exhaustion series and the product-to-waste concentration ratio study, the membranes were examined for deterioration.

During the second exhaustion series and the product-to-waste ratio study, the stack electrical resistance was consistently higher in System I than in System II. The electrical resistance pattern of System I, moreover, was quite erratic, apparently as a result of the build-up and periodic sloughing of slime or organic deposits from the dilute stream side of the anion membranes. The generally higher electrical resistance of System I was due both to this organic deposition and to the sorption of organic ions by the anion membranes. Examination of the membranes in System I at the end of the test period showed an increase in anion membrane resistivities and a decrease in anion membrane capacities. Elution of ABS from these membranes reduced the resistivities nearly to their original values and partially restored the capacities.

Anion membranes from System II were not particularly affected by organics; no cation membranes from either system showed any signs of deterioration. An increase in the waste-to-product stream concentration ratio above about 40 to 1 resulted in a noticeable increase in stack electrical resistance; the effect, however, was not as great as the effects of organic deposition, scaling, or membrane fouling.

Current efficiencies were reduced with increasing waste-to-product concentration ratios, due to increasing back-diffusion. System II current efficiencies were nearly always lower than those of System I; however, this was suspected to be more a function of the method of calculation rather than due to any differences between the systems. Energy consumption calculations were similar for both systems, although the use of "actual" current efficiencies would probably result in much increased energy consumptions for System I. Beyond a waste-to-product concentration ratio of about 50 to 1, energy consumptions increased sharply.

The importance of fairly complete organic removal prior to electrodialysis was shown throughout the exhaustion testing. Slime or organic deposits on the membrane surfaces were shown to result in a considerable reduction of allowable current densities due to increased boundary film thickness.

ABS concentrations in the influent to the System II electrodialysis stack averaged less than 0.10 milligram per liter, and no particular capacity loss or resistivity gain was noted at the end of the exhaustion testing in any of the membranes. With an influent COD of about 20 to 40 milligrams per liter, however, light to moderate organic deposits were found on the anion membranes after 650 hours of operation.

To ensure minimal maintenance and the most efficient operation of an electrodialysis system in a waste-water renovation application, the electrodialysis stack(s) should logically be placed at the end of a chain of processes where the waste water would have its lowest concentration of organic material and highest concentration of inorganics.

The effect of prior disinfection of the feed upon organic deposits or slime growths within an electrodialysis stack has not yet been investigated.

B. SELECTIVITY

The transport of ions across an ion-exchange membrane has been discussed in terms of chemical and electrical potential gradients and convection. The interdependency of the transport of each ionic species upon all other species was described by a system of simultaneous equations. Relative rates of removal of individual ionic species are represented by a separation factor:

$$\alpha^{i}_{k} = \frac{dC_i/C_i}{dC_k/C_k} \qquad (25)$$

The effects of solution pH and mole fraction upon a phosphate - chloride separation factor were investigated by means of a modified electrodialysis stack. A constant total solution concentration of 5 millimoles per liter, flow path velocity of 6.4 centimeters per second, and current of 20 milliamperes (effective membrane area = 212 cm^2 per cell pair) were used. At pH 4.5, the phosphate - chloride separation factor increased from 0.4 at very high phosphate mole fractions to 0.5 at very low phosphate mole fractions. At pH 8.5, the phosphate - chloride separation factor increased from 0.6 to 1.0 with decreasing phosphate mole fraction. The test membrane was the same type used in the exhaustion tests, a Nepton AR111BZL-183.

The effects of other solution parameters, e.g., total concentration, upon the phosphate - chloride and other separation factors have not yet been investigated.

A method for qualitative estimation of the effects of membrane parameters upon selectivity was proposed, which neglects the contribution of a diffusion gradient to ionic transport across ion-exchange membranes. Separation factors were estimated from static equilibrium data. Four different membrane types (all standard anion membranes) were investigated. Nitrate, sulfate, and mono-, di-, and tri-valent phosphate removal estimates were related to estimated chloride removals by separation factors and flux ratios.

Although these estimates must be verified by operation of these membrane types in a test electrodialysis stack, the data obtained indicate that an increase in membrane porosity increases the rates of removal of nitrate, sulfate, and phosphate relative to chloride removals.

As in the solution parameter study, phosphate removals increased with pH in the membrane parameter study, also.

Analytical data from all exhaustion runs and the product-to-waste ratio studies were plotted in the form of percent species removed versus percent total equivalents removed for all cations and anions except phosphate; phosphate removal data were plotted as phosphate separation factor versus time of operation. Although this removal data encompasses the entire test period and all the variations experienced in solution concentrations, temperatures, pH's, operating conditions, degrees of membrane fouling, scaling, and sliming, nevertheless reasonable estimates of separation factors were obtained for most ionic species as functions of certain operating conditions.

No noticeable difference in cationic separation factors was observed between the two systems, or between the exhaustion series and the product-to-waste ratio studies.

Sodium separation factors were nearly always very close to 1.0.

Total hardness separation factors varied between 2.5 for membrane-controlled transfer and 1.2 for diffusion-controlled transfer.

Ammonium removals were most erratic, probably because of the large variation in ammonium concentrations in the sewage effluent. Ammonium transfer will require further study.

Anionic separation factors show the effects of anion membrane fouling, total cut, and concentrate-stream acid feed upon relative rates of removal.

Chloride removal was enhanced by membrane fouling (other species' removals were depressed). Since fouling tends to clog membrane pores, which is similar to reducing membrane porosity, this would increase chloride transfer at the expense of other anionic species. Chloride separation factors were depressed with increasing total cut or approach to film-diffusion limiting transfer. The use of HCl as concentrate-stream acid feed depressed the chloride separation factors still further because of the large back-diffusion gradient. Film-diffusion limited chloride separation factors with HCl feed to concentrate streams approached 0.8 for both systems.

Bicarbonate separation factors were depressed by membrane fouling, increased with approach to film-diffusion limiting transfer, and increased still further with concentrate-stream acid treatment. Film-diffusion controlled bicarbonate separation factors with HCl feed to concentrate stream approached 1.6 for System I and 2.0 for System II.

Nitrate separation factors were erratic, probably due to extreme variations in nitrate concentrations. Nitrate removals behaved somewhat the same as chloride removals, i.e., they increased with membrane fouling. Concentrate-stream acid treatment did not noticeably affect nitrate separation factors. Film-diffusion limiting separation factors approached 1.5 for System I and slightly more than 1.0 for System II.

Sulfate separation factors, also, were erratic. The data indicate that sulfate removal is depressed by membrane fouling. Film-diffusion limiting separation factors for the "clean" System II were slightly greater than 1.0. All separation factors for System I were less than 1.0.

Plotted against operating time, phosphate separation factors were clearly depressed with membrane fouling. Separation factors greater than 2.0 were realized with "clean" stacks. With very fouled stacks, phosphate removals were reduced below the level of analytical detection.

The selectivity studies were not designed necessarily to produce comprehensive quantitative results. Rather, their purpose was to illustrate the dependency of the various separation factors upon a variety of parameters and to indicate where more intensive study might be warranted.

C. PRELIMINARY DESIGN AND COST ESTIMATE OF A 10-MGD ELECTRODIALYSIS PLANT

The necessary electrodialysis design equations were extracted from the literature and modified for waste-water renovation application. Design criteria were developed from United States Public Health Service data and tentative specifications and from the results of research conducted to date under Contract No. SAph 76690. A preliminary design was then described, which resulted in the selection of 44 Ionics' Mark III, 275 cell-pair, electrodialysis stacks and necessary appurtenances. Forty-two of these stacks operated in parallel will reduce the concentration of a 10-MGD product stream from 850 to 500 milligrams per liter at a cost of 7.39 cents per thousand gallons of product water.

SECTION VI
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APPENDIX A

Systems Operation

TABLE A-1.

SYSTEMS OPERATION: EXHAUSTION RUNS - SERIES I

Date/Time	System I	Hours of Operation		System II
1962 Dec. 11 10:00 AM	Rebuilt stack placed back in operation E = 5V I = 50 ma Flow rate = 130 ml/min	0		
Dec. 13 1:30 PM	E = 5V I = 45 ma Flow rate = 132 ml/min	51.5	0	Started System II
Dec. 23 10:00 AM	E = 5V I = 26 ma Flow rate = 125 ml/min Dilute R_{in} = 5,200 ohm-cm R_{out} = 6,100 ohm-cm	288		E = 5V I = 65 ma Flow rate = 136 ml/min Dilute R_{in} = 5,200 ohm-cm R_{out} = 6,700 ohm-cm
1:30 PM	2 liters 0.5N HCl pumped through dilute stream		240	
2:00 PM	E = 5V I = 55 ma	292		
Dec. 24 4:00 PM	Shut down	318	266	Shut down
Dec. 26 9:00 AM	Started system	318	266	Started system
11:30 AM	E = 5V I = 27 ma			E = 5V I = 40 ma
11:45 AM	E = 10V I = 54 ma			
Dec. 27 3:00 PM			296	Shut down for charcoal replacement

TABLE A-1 (continued)

SYSTEMS OPERATION: EXHAUSTION RUNS - SERIES 1

Date/Time	System I	Hours of Operation		System II
Dec. 29				
12:30 PM	E = 7.5V I = 52 ma Received new load of sewage from Clinton	393		
1:30 PM	E = 10V I = 50 ma			
Dec. 30				
Noon	E = 10V I = 50 ma Flow rate = 120 ml/min Voltage drops through stack measured with probes	417		
3:30 PM	1.2 liter 1N HCl pumped through conc. and electr. streams			
5:00 PM	Stack again probed for voltage drops			
Dec. 31				
Noon		441	296	Charcoal replaced. Started System E = 25V I = 30 ma
6:00 PM	E = 25V I = 98 ma Dilute R_{out} = 12,500 ohm-cm System shut down	447	302	System shut down
1963				
Jan. 2				
9:30 AM	Started system	447	302	Started system
10:45 AM	E = 25V I = 116 ma			E = 25V I = 130 ma
12:30 PM	Probed stack			Probed stack
8:00 PM			313	Reduced voltage E = 12.5V I = 105 ma Changed charcoal because of cloudiness in reservoir
11:00 PM				E = 15V I = 100 ma
Jan. 3				
9:30 AM	E = 23.8V I = 173 ma Dilute R_{in} = 2940 ohm-cm R_{out} = 5150 ohm-cm	471	326	E = 13.8V I = 190 ma Dilute R_{in} = 2950 ohm-cm R_{out} = 5500 ohm-cm

TABLE A-1 (continued)

SYSTEMS OPERATION: EXHAUSTION RUNS - SERIES I

Date/Time	System I	Hours of Operation		System II
Jan 3 cont'd				
3:45 PM	E = 24.2V I = 205 ma Dilute R_{out} = 5150 ohm-cm	477	332	E = 14V I = 170 ma Dilute R_{out} = 10,300 ohm-cm Conc stream plugged 65 ml conc. HCl poured in 4 liter conc. reservoir
4:00 PM			333	E = 5V I = 110 ma Dilute R_{out} = 4,780 ohm-cm
4:30 PM	E = 15V I = 120 ma Added 60 ml conc. HCl to 4 liter conc. reservoir E = 7V I = 130 ma Dilute R_{out} = 3280 ohm-cm			
5:00 PM	E = 10V I = 116 ma Dilute R_{out} = 3730 ohm-cm			
6:00 PM				Changed charcoal
Jan. 4				
2:00 PM	E = 18V I = 116 ma Dilute R_{out} = 3350 ohm-cm			E = 6V I = 115 ma Dilute R_{out} = 3660 ohm-cm
4:30 PM	E = 19V I = 105 ma		357	E = 7V I = 115 ma
5:00 PM	Shut down	502		Shut down
Jan. 5	Disassembled stack; examined membranes			Disassembled stack; examined membranes

TABLE A-2.

SYSTEMS OPERATION: EXHAUSTION RUNS - SERIES 2

Date/Time	System I	Hours of Operation		System II
1963				
Jan. 28				
2.00 PM			0	Started operation
4:30 PM			2.5	E = 15V I = 100 ma
5:00 PM	Started operation	0		
6:10 PM	E = 15V I = 100 ma	1		
Jan. 29				
10:30 AM	E = 15V I = 100 ma			E = 16V I = 100 ma
11:00 AM	Shut down because sewage had run out	18	21	Shut down
Jan. 30				
11:00 AM	New load of sewage arrived Started system E = 15V I = 116 ma R_{out} = 18,400 ohm-cm Temp. = 13°C	18	21	Started system E = 15V I = 120 ma R_{out} = 9900 ohm-cm Temp. = 14°C
7:30 PM	E = 15V I = 130 ma R_{out} = 9150 ohm-cm Temp. = 14°C	26.5	29.5	E = 15V I = 150 ma R_{out} = 6830 ohm-cm Temp. = 16°C
Jan. 31				
8:30 AM	Found both systems empty Shut down system.	39.5	42.5	Shut down system
	Conductivity bridge checked and found to be unreliable. Replaced with new bridge. Found source of system drainage and corrected.			
Feb. 4				
11:00 AM	Started system E = 15V I = 104 ma R_{out} = 22,500 ohm-cm Temp. = 13°C Flow rate = 110 ml/min	39.5	42.5	Started system E = 15V I = 100 ma R_{out} = 9500 ohm-cm Temp. = 13°C Flow rate = 120 ml/min

TABLE A-2 (continued)

SYSTEMS OPERATION: EXHAUSTION RUNS - SERIES 2

Date/Time	System I	Hours of Operation		System II
Feb. 4 cont'd				
6:30 PM	E = 16V I = 102 ma R_{out} = 15,000 ohm-cm Temp. = 17°C	47	50	E = 15V I = 104 ma R_{out} = 7900 ohm-cm Temp. = 16°C
Feb. 5				
9:45 AM	E = 15V I = 80 ma R_{out} = 10,400 ohm-cm Temp. = 15°C pH = 6.4	62	65	E = 15V I = 90 ma R_{out} = 7500 ohm-cm Temp. = 17°C pH = 7.0
5:30 PM	E = 16V I = 80 ma R_{out} = 9500 ohm-cm Temp. = 17°C pH = 6.35	70	73	E = 15V I = 108 ma R_{out} = 6600 ohm-cm Temp. = 18°C pH = 6.7
7:30 PM	Shut down	72	75	Shut down
Feb. 6				
8:15 AM	Started system	72	75	Started system
9:30 AM	E = 15V I = 94 ma R_{out} = 15,000 ohm-cm Temp. = 16°C pH = 5.8	73	76	E = 15V I = 112 ma R_{out} = 6700 ohm-cm Temp. = 18.5°C pH = 6.65
4:00 PM	E = 15.1V I = 95 ma R_{out} = 7850 ohm-cm Temp. = 17°C pH = 6.45	80	83	E = 15V I = 108 ma R_{out} = 6300 ohm-cm Temp. = 19.5°C pH = 6.75
4:45 PM	Shut down	84	87	Shut down
Feb. 7				
10:15 AM	Started pumping. Reversed current for one minute E = 10V I = 230 ma			Started pumping. Reversed current for one minute E = 10V I = 150 ma
10:20 AM	Normal operation	84	87	Normal operation
10:50 AM	E = 15V I = 91 ma R_{out} = 13,900 ohm-cm Temp. = 14°C pH = 6.1 Q = 106 ml/min			E = 15V I = 100 ma R_{out} = 9100 ohm-cm Temp. = 14°C pH = 6.5 Q = 122 ml/min
11:00 AM	Increased pump setting to 48.5. Q = 116 ml/min.	84.5	87.5	
Noon	New sewage. Added 20 gm $NaHO_3$ to 200 gal.			
4:00 PM	E = 15V I = 137 ma R_{out} = 4450 ohm-cm Temp. = 15.3°C pH = 6.7 Q = 117 ml/min	89.5	92.5	E = 15V I = 110 ma R_{out} = 3230 ohm-cm Temp. = 15.4°C pH = 8.6 Q = 123 ml/min

TABLE A-2 (continued)

SYSTEMS OPERATION: EXHAUSTION RUNS - SERIES 2

Date/Time	System I	Hours of Operation		System II
Feb. 7 cont'd				
4:15 PM	E = 12V I = 107.5 ma pH = 7.0			
5:30 PM	E = 10.5V I = 109 ma			Sewage very dark, cloudy green. Charcoal not helping much.
6:00 PM	Increased pump setting to 49. Q = 121 ml/min	91.5	94.5	
Feb. 8				
9:45 AM	E = 10.5V I = 75 ma R_{out} = 3750 ohm-cm Temp. = 10°C pH = 7.15	107	110	E = 14.5V I = 90 ma R_{out} = 3950 ohm-cm Temp. = 10°C pH = 7.05
10:00 AM	E = 15V I = 104 ma			E = 15V I = 95 ma
4:30 PM	Power failure. All equipment shut off.	114	117	Shut down
Feb. 9	Power restored. Drained sewage from tank because it appeared septic. New load arrived from Clinton. Added 20 gm $NaNO_3$			
Feb. 11				
11:00 AM	Started operation E = 15V I = 149 ma R_{out} = 6500 ohm-cm Temp. = 18°C pH = 6.1	114	117	Started operation E = 15V I = 155 ma R_{out} = 6500 ohm-cm Temp. = 22°C pH = 7.12
1:45 PM	E = 16V I = 90 ma R_{out} = 9900 ohm-cm Temp. = 18°C pH = 7.1 R_{in} = 3500 ohm-cm T_{in} = 17.5°C	117	120	E = 15V I = 65 ma R_{out} = 5600 ohm-cm Temp. = 18.5°C pH = 7.6 R_{in} = 3550 ohm-cm T_{in} = 20°C
2:30 PM	Voltage probe taken	117.5	120.5	Voltage probe taken
7:15 PM	E = 17.5V I = 83 ma	122	125	E = 16.2V I = 58 ma
Feb. 12				
9:45 AM	E = 16V I = 76 ma R_{out} = 7800 ohm-cm Temp. = 16°C pH = 6.65 Q_{out} = 115 ml/min	137	140	E = 15V I = 59 ma R_{out} = 5000 ohm-cm Temp. = 16°C pH = 7.6 Q_{out} = 116 ml/min
10:00 AM	E = 13V I = 63 ma			

TABLE A-2 (continued)

SYSTEMS OPERATION: EXHAUSTION RUNS - SERIES 2

Date/Time	System I	Hours of Operation		System II
Feb. 12 cont'd				
6:15 PM	E = 13V I = 66 ma R_{out} = 6300 ohm-cm T = 14°C pH = 6.75	145	148	E = 15V I = 59 ma R_{out} = 5000 ohm-cm T = 14°C pH = 7.1
Feb. 13		159	162	
8:15 AM	Reservoirs empty. Shut down systems.			
Feb. 14				
11:00 AM	Started operation	159	162	Started operation
2:00 PM	E = 10.5V I = 86 ma R_{out} = 3570 ohm-cm T = 11.3°C pH = 7.1 Increased E to 13V I to 100 ma	162	165	E = 10.5V I = 48 ma R_{out} = 3000 ohm-cm T = 12.5°C pH = 8.9 Increased E to 15V I to 100 ma
6:00 PM	E = 14V I = 95 ma	166	169	E = 16V I = 85 ma
Feb. 15				
3:30 PM	E = 13.5V I = 70 ma R_{out} = 4500 ohm-cm T = 11.5°C pH = 7.1 Q = 116 ml/min	187.5	190.5	E = 15V I = 60 ma R_{out} = 3900 ohm-cm T = 10.8°C pH = 7.21 Q = 125 ml/min
Feb. 16				
12:30 PM	Reservoirs empty. Shut down systems.	208.5	211.5	
Feb. 18				
11:00 AM	Started operation	208.5	211.5	Started operation
2:00 PM	E = 15V I = 93 ma R_{out} = 5780 ohm-cm T = 17°C pH = 6.21 Q = 120 ml/min	211.5	214.5	E = 15V I = 60 ma R_{out} = 5000 ohm-cm T = 17°C pH = 9.0 Q = 120 ml/min
3:00 PM	E = 10V I = 60 ma Voltage probe taken	212.5	215.5	Voltage probe taken
3:50 PM	E = 10V I = 60 ma R_{out} = 5300 ohm-cm T = 15°C pH = 7.0	213.5		

TABLE A-2 (continued)

SYSTEMS OPERATION: EXHAUSTION RUNS - SERIES 2

Date/Time	System I	Hours of Operation		System II
Feb. 19				
7:45 AM	E = 10V I = 60 ma R_{out} = 4200 ohm-cm T = 20°C pH = 6.80 Q = 120 ml/min R_{in} = 3100 ohm-cm T_{in} = 18°C pH_{in} = 7.2	229.5	232.5	E = 15V I = 50 ma R_{out} = 3800 ohm-cm T = 19°C pH = 7.2 Q = 120 ml/min R_{in} = 3100 ohm-cm T_{in} = 18°C pH_{in} = 7.35
8:00 AM to Noon	Polarization study			Polarization study
4:00 PM	E = 11.8V I = 59 ma R_{out} = 5200 ohm-cm T = 19°C pH = 7.2 Q = 120 ml/min R_{in} = 3300 ohm-cm T_{in} = 18°C pH_{in} = 7.5	237.5	240.5	E = 14.1V I = 51 ma R_{out} = 3800 ohm-cm T = 18°C pH = 7.3 Q = 120 ml/min R_{in} = 3050 ohm-cm T_{in} = 18°C pH_{in} = 7.5
Feb. 20				
8:00 AM	E = 10V I = 42 ma R_{out} = 4600 ohm-cm T = 16°C pH = 7.0 Q = 115 ml/min R_{in} = 3000ohm-cm T_{in} = 19°C pH_{in} = 7.15 Voltage probe taken	253.5	256.5	E = 13V I = 52 ma R_{out} = 3750 ohm-cm T = 18°C pH = 7.3 Q = 130 ml/min R_{in} = 3000 ohm-cm T_{in} = 21°C pH_{in} = 7.3 Voltage probe taken
12:20 PM	E = 10V I = 40 ma			E = 13.5V I = 55 ma
			261	Shut down System II
12:55 PM	Shut down System I	258.5		
Feb. 21				
10:50 AM	Started operation	258.5	261	Started operation
11:00 AM	E = 4.3V I = 42 ma R_{out} = 3300 ohm-cm T = 14°C pH = 6.92 Q = 115 ml/min R_{in} = 2650 ohm-cm T_{in} = 15.5°C pH_{in} = 6.92			E = 11V I = 50 ma R_{out} = 3300 ohm-cm T = 16°C pH = 7.08 Q = 122.5 ml/min R_{in} = 2650 ohm-cm T_{in} = 16°C pH_{in} = 7.08
7:30 PM	E = 5V I = 30 ma	267	269.5	E = 12.8V I = 42 ma
Feb. 22				
1:00 PM	E = 5V I = 28 ma R_{out} = 3800 ohm-cm T = 18°C pH = 7.0	284.5	287	E = 12.5V I = 45 ma R_{out} = 3100 ohm-cm T = 15°C pH = 7.2
1:15 PM	Sewage very dark. Septic smelling. Shut down both systems.			

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TABLE A-2 (continued)

SYSTEMS OPERATION: EXHAUSTION RUNS - SERIES 2

Date/Time	System I	Hours of Operation		System II
Feb. 25				
1:00 PM	Started operation	285	287	Started operation
2:15 PM	E = 5V I = 32 ma R_{out} = 4200 ohm-cm T = 12°C pH = 8.25 Q = 120 ml/min R_{in} = 2900 ohm-cm T_{in} = 15°C pH_{in} = 7.75	286	288	E = 10V I = 30 ma R_{out} = 3900 ohm-cm T = 13°C pH = 9.75 Q = 125 ml/min R_{in} = 3700 ohm-cm T_{in} = 14°C pH_{in} = 10
Feb. 26				
8:00 AM	E = 5V I = 29 ma R_{out} = 3900 ohm-cm T = 15°C pH = 7.15 Q = 115 ml/min R_{in} = 2970 ohm-cm T_{in} = 16°C pH_{in} = 7.22	305	307	E = 12V I = 22 ma R_{out} = 3400 ohm-cm T = 15°C pH = 7.90 Q = 125 ml/min R_{in} = 3050 ml/min T_{in} = 15°C pH_{in} = 8.50
8:30 AM	Voltage probe taken			Voltage probe taken
3:30 PM	E = 5.2V I = 28 ma R_{out} = 3800 ohm-cm T = 16°C pH = 7.1 R_{in} = 2800 ohm-cm T_{in} = 17°C pH_{in} = 7.1	312.5	314.5	E = 11.5V I = 22 ma R_{out} = 3250 ohm-cm T = 18°C pH = 7.1 R_{in} = 2800 ohm-cm T_{in} = 17°C pH_{in} = 7.38
Feb. 27				
8:00 AM	E = 5V I = 24 ma R_{out} = 3700 ohm-cm T = 14.5°C pH = 7.1 Q = 112.5 ml/min R_{in} = 2900 ohm-cm T_{in} = 17°C pH_{in} = 7.2	329	331	E = 11.5V I = 23 ma R_{out} = 3400 ohm-cm T = 16°C pH = 7.35 Q = 125 ml/min R_{in} = 3250 ohm-cm T_{in} = 17°C pH_{in} = 7.4
4:30 PM	Voltage probe taken	337.5	339.5	Voltage probe taken
5:30 PM	Shut down	338.5	340.5	Shut down
Mar. 1				Dismantled stack. $CaCO_3$ scaling found in conc. streams. Removed with HCl and reassembled stack.
Mar. 2				
12:15 PM	Started ~peration	338.5	340.5	Started operation
12:35 PM	E = 3.9V I = 30 ma			E = 2.5V I = 32 ma

TABLE A-2 (continued)

SYSTEMS OPERATION: EXHAUSTION RUNS - SERIES 2

Date/Time	System I	Hours of Operation		System II
Mar. 2 cont'd				
2:00 PM				Added 25 ml HCl (conc.) to conc. stream
5:30 PM	E = 4V I = 20.5 ma	344	346	E = 2.6V I = 23 ma
Mar. 3				
1:45 PM	E = 4.1V I = 19 ma			E = 2.6V I = 24 ma
6:30 PM	E = 4.1V I = 19 ma	369	371	E = 2.6V I = 24 ma
Mar. 4				
8:00 AM	Shut down Reservoir empty	382.5	384.5	Shut down Reservoir empty
Mar. 7				
8:30 AM	Started operation	382.5	384.5	Started operation
10:15 AM	E = 6V I = 38 ma R_{out} = 4700 ohm-cm T = 14.6°C pH = 7.32 Q = 115 ml/min R_{in} = 3600 ohm-cm T_{in} = 12.5°C pH_{in} = 7.7			E = 3.1V I = 33 ma R_{out} = 4600 ohm-cm T = 15.5°C pH = 9.0 Q = 120 ml/min R_{in} = 3700 ohm-cm T_{in} = 16.5°C pH_{in} = 9.15
Noon	Voltage probe taken	386	388	Voltage probe taken
Mar. 8				
8:15 AM	E = 7V I = 30 ma R_{out} = 4400 ohm-cm T = 13°C pH = 7 Q = 115 ml/min R_{in} = 3300 ohm-cm T_{in} = 14°C pH_{in} = 7 Voltage probe taken	406	408	E = 3.5V I = 34 ma R_{out} = 4200 ohm-cm T = 13.5°C pH = 7 Q = 120 ml/min R_{in} = 3300 ohm-cm T_{in} = 14°C pH_{in} = 7.35 Voltage probe taken
4:30 PM	E = 7V I = 38 ma R_{out} = 4300 ohm-cm T = 15°C pH = 6.8 R_{in} = 3300 ohm-cm T_{in} = 16°C pH_{in} = 7	414	416	E = 3.5V I = 36 ma R_{out} = 4000 ohm-cm T = 15°C pH = 7 R_{in} = 3300 ohm-cm T_{in} = 15°C pH_{in} = 7.2
Mar. 9				
10:30 AM			434	Shut down
11:00 AM	Shut down	432.5		

TABLE A-2 (continued)

SYSTEMS OPERATION: EXHAUSTION RUNS - SERIES 2

Date/Time	System I	Hours of Operation		System II
Mar. 11				
2:00 PM	Started system E = 4V I = 30 ma R_{out} = 3550 ohm-cm T = 13.5°C pH = 6.9 R_{in} = 2700 ohm-cm T_{in} = 16.4°C pH_{in} = 6.6 Q_{out} = 120 ml/min Voltage probe taken	432.5	434	Started system E = 3.5V I = 38 ma R_{out} = 3520 ohm-cm T = 13.5°C pH = 8.5 R_{in} = 2700 ohm-cm T_{in} = 16.5°C pH_{in} = 8.3 Q_{out} = 123 ml/min Voltage probe taken
Mar. 12				
9:00 AM	E = 5V I = 20 ma R_{out} = 4800 ohm-cm T = 24.5°C pH = 7 R_{in} = 3000 ohm-cm T_{in} = 18.5°C pH_{in} = 7 Q_{out} = 113 ml/min	451.5	453	E = 3V I = 22 ma R_{out} = 5200 ohm-cm T = 15.5°C pH = 7.2 R_{in} = 3500 ohm-cm T_{in} = 18°C pH_{in} = 7.2 Q_{out} = 135 ml/min
9:30 AM	Voltage probe taken	452	453.5	Voltage probe taken
4:00 PM	E = 5.5V I = 22 ma R_{out} = 4800 ohm-cm T = 16.5°C pH = 7 R_{in} = 3500 ohm-cm T_{in} = 18°C pH_{in} = 7	458.5	460	E = 3.2V I = 32 ma R_{out} = 4800 ohm-cm T = 16.5°C pH = 7.15 R_{in} = 3400 ohm-cm T_{in} = 19°C pH_{in} = 7.15
Mar. 13				
2:00 PM	E = 5V I = 22 ma R_{out} = 4500 ohm-cm T = 16.5°C pH = 6.8 R_{in} = 3500 ohm-cm T_{in} = 19°C pH_{in} = 7.08 Voltage probe taken	480.5	482	E = 3V I = 31 ma R_{out} = 4700 ohm-cm T = 15°C pH = 7 R_{in} = 3550 ohm-cm T_{in} = 18°C pH_{in} = 7 Voltage probe taken
3:00 PM	E = 7.8V I = 30 ma R_{out} = 5000 ohm-cm T = 18°C pH = 6.9 R_{in} = 3500 ohm-cm T_{in} = 19°C pH_{in} = 7	481.5	483	E = 4.6V I = 40 ma R_{out} = 5000 ohm-cm T = 18°C pH = 6.9 R_{in} = 3550 ohm-cm T_{in} = 18°C pH_{in} = 7
4:00 PM	Shut down system	482.5	484	Shut down system
Mar. 14				
Noon	Started system	482.5	484	Started system

TABLE A-2 (continued)

SYSTEMS OPERATION: EXHAUSTION RUNS - SERIES 2

Date/Time	System I	Hours of Operation		System II
Mar. 14 cont'd 3:00 PM	E = 6V I = 22 ma R_{out} = 4900 ohm-cm T = 15°C pH = 7.18 R_{in} = 4000 ohm-cm T_{in} = 16°C pH_{in} = 7.25 Q_{out} = 115 ml/min	485.5	487	E = 3V I = 28 ma R_{out} = 4900 ohm-cm T = 15.5°C pH = 8.65 R_{in} = 4000 ohm-cm T_{in} = 16°C pH_{in} = 8.6 Q_{out} = 120 ml/min
Mar. 15 3:30 PM	E = 5V I = 20 ma R_{out} = 4300 ohm-cm T = 15.5°C pH = 6.9 R_{in} = 3300 ohm-cm T_{in} = 17.5°C pH_{in} = 7.1 Q_{out} = 105 ml/min	510	511.5	E = 3V I = 25 ma R_{out} = 4200 ohm-cm T = 15°C pH = 7.5 R_{in} = 3300 ohm-cm T_{in} = 18.5°C pH_{in} = 7.4 Q_{out} = 125 ml/min
Mar. 16 2:00 PM	Shut down system	532.5	534	Shut down system
Mar. 18 1:00 PM	Started system Conducted polarization run	532.5	534	Started system Conducted polarization run
Mar. 19 8:00 AM	E = 4.9V I = 19 ma R_{out} = 4600 ohm-cm T = 13°C pH = 7.4 R_{in} = 3000 ohm-cm T_{in} = 17.4°C pH_{in} = 7.4	551.5	553	E = 3V I = 30 ma R_{out} = 4000 ohm-cm T = 12.5°C pH = 7.4 R_{in} = 3100 ohm-cm T_{in} = 17°C pH_{in} = 7.5
Mar. 20 8:00 AM	E = 6V I = 30 ma R_{out} = 4300 ohm-cm T = 15.8°C pH = 7.4 R_{in} = 3200 ohm-cm T_{in} = 16.6°C pH_{in} = 7.4 Q_{out} = 102 ml/min	575.5	577	E = 2.8V I = 20 ma R_{out} = 3700 ohm-cm T = 16°C pH = 7.4 R_{in} = 3150 ohm-cm T_{in} = 19°C pH_{in} = 7.35 Q_{out} = 125 ml/min
9:00 AM	Voltage probe taken	576.5	578	Voltage probe taken
1:15 PM			582	Shut down system
2:10 PM	Shut down system	581.5		

TABLE A-2 (continued)

SYSTEMS OPERATION: EXHAUSTION RUNS - SERIES 2

Date/Time	System I	Hours of Operation		System II
Mar. 21				
1:30 PM	Started system	581.5	582	Started system Polarization run
Mar. 22				
8:00 AM	E = 4.5V I = 50 ma R_{out} = 4100 ohm-cm T = 16°C pH = 6.80 R_{in} = 2900 ohm-cm T_{in} = 17°C pH_{in} = 7.05	600	600.5	E = 5.0V I = 60 ma R_{out} = 4300 ohm-cm T = 16.3°C pH = 6.25 pH_{in} = 6.6 Polarization run
2:00 PM	Dismantled System I. Found scaling on conc. side of anion membranes. Slime on dilute side. Washed with dilute HCl.	606		
3:15 PM	Reassembled System I.			
3:30 PM	Started operation Polarization run	606		
4:30 PM	E = 5V I = 50 ma	607		E = 5.5V I = 60 ma
Mar. 23				
2:00 PM	Shut down system	628.5	630.5	Shut down system
Mar. 25				
8:00 AM	Started system	628.5	630.5	Started system
11:15 AM	E = 4V I = 27 ma R_{out} = 6000 ohm-cm T = 17.6°C pH = 7 R_{in} = 3100 ohm-cm T_{in} = 20.5°C pH_{in} = 7.2 Q_{out} = 107 ml/min Voltage probe taken E = 3.9V I = 29 ma R_{out} = 6200 ohm-cm T = 15.5°C pH = 6.5 pH_{in} = 6.6	631.5	633.5	E = 4V I = 27 ma R_{out} = 6800 ohm-cm pH = 7 R_{in} = 5200 ohm-cm T_{in} = 18.5°C pH_{in} = 7 Q_{out} = 120 ml/min Voltage probe taken E = 4V I = 28 ma R_{out} = 6800 ohm-cm T = 16°C pH = 6.6 pH_{in} = 6.6

TABLE A-2 (continued)

SYSTEMS OPERATION: EXHAUSTION RUNS - SERIES 2

Date/Time	System I	Hours of Operation		System II
Mar. 26				
10:00 AM	$E = 3.9V$ $I = 32$ ma $R_{out} = 7000$ ohm-cm $T = 18°C$ $pH = 6.7$ $R_{in} = 3600$ ohm-cm $T_{in} = 21.5°C$ $pH_{in} = 6.83$	654.5	656.5	$E = 4V$ $I = 31$ ma $R_{out} = 5200$ ohm-cm $T = 17.5°C$ $pH = 6.85$ $R_{in} = 3500$ ohm-cm $T_{in} = 22.5°C$ $pH_{in} = 6.85$
11:00 AM	Voltage probe taken	655.5	657.5	Voltage probe taken
4:00 PM	$E = 3.9V$ $I = 32$ ma $R_{out} = 6400$ $pH = 6.85$ $R_{in} = 3600$ ohm-cm $T_{in} = 17°C$ $pH_{in} = 7$	660.5	662.5	$E = 4V$ $I = 31$ ma $R_{out} = 6000$ $pH = 6.9$ $R_{in} = 3600$ ohm-cm $T_{in} = 22°C$ $pH_{in} = 6.85$
Mar. 27				
8:00 AM	Shut down system	676.5	678.5	Shut down system
Mar. 28				
9:45 AM	Started system $E = 4.3V$ $I = 33$ ma $R_{out} = 8500$ ohm-cm $T = 14.5°C$ $pH = 7$ $R_{in} = 5000$ ohm-cm $T_{in} = 14.5°C$ $pH_{in} = 7$ $Q_{out} = 110$ ml/min	676.5	678.5	Started system $E = 4.8V$ $I = 28$ ma $R_{out} = 8000$ ohm-cm $T = 18°C$ $pH = 8.3$ $R_{in} = 6800$ ohm-cm $T_{in} = 17.3°C$ $pH_{in} = 9$ $Q_{out} = 125$ ml/min
Mar. 29				
8:00 AM	$E = 4.5V$ $I = 33$ ma $R_{out} = 7600$ ohm-cm $T = 14.5°C$ $pH = 7.2$ $R_{in} = 4500$ ohm-cm $T_{in} = 18.5°C$ $pH_{in} = 7.1$	698.5	700.5	$E = 5V$ $I = 37$ ma $R_{out} = 6000$ ohm-cm $T = 16°C$ $pH = 7.35$ $R_{in} = 4000$ ohm-cm $T_{in} = 18.5°C$ $pH_{in} = 7.38$
3:30 PM	$E = 4.8V$ $I = 34$ ma $R_{out} = 7800$ ohm-cm $T = 15.4°C$ $pH = 6.5$ $R_{in} = 4000$ ohm-cm $T_{in} = 17.5°C$ $pH_{in} = 6.7$ $Q_{out} = 105$ ml/min Voltage probe taken	706	708	$E = 5V$ $I = 39$ ma $R_{out} = 6200$ ohm-cm $T = 15.5°C$ $pH = 6.7$ $R_{in} = 4400$ ohm-cm $T_{in} = 16.8°C$ $pH_{in} = 6.7$ $Q_{out} = 125$ ml/min Voltage probe taken
Mar. 30				
1:15 PM	Shut down system	727	729	Shut down system

TABLE A-2 (continued)

SYSTEMS OPERATION: EXHAUSTION RUNS - SERIES 2

Date/Time	System I	Hours of Operation		System II
Apr. 1				
10:00 AM	Started system	727	729	Started system
11:00 AM	E = 3.8V I = 48 ma R_{out} = 2400 ohm-cm T = 17.4°C pH = 7.1 R_{in} = 1700 ohm-cm T_{in} = 17°C pH_{in} = 7.35 Q_{out} = 105 ml/min Voltage probe taken	728	730	E = 4V I = 49 ma R_{out} = 2700 ohm-cm T = 18°C pH = 7.4 R_{in} = 1750 ohm-cm T_{in} = 18°C pH_{in} = 7.49 Q_{out} = 125 ml/min Voltage probe taken
4:00 PM	E = 5V I = 47 ma R_{out} = 7000 ohm-cm T = 16°C pH = 6.7 R_{in} = 3500 ohm-cm pH_{in} = 6.7	733	735	E = 4.8V I = 49 ma R_{out} = 4700 ohm-cm T = 16°C pH = 7 R_{in} = 3500 ohm-cm pH_{in} = 6.6
Apr. 2				
3:15 PM	E = 6.5V I = 37 ma R_{out} = 8000 ohm-cm T = 18°C pH = 6.65 R_{in} = 4000 ohm-cm T_{in} = 18°C pH_{in} = 6.85 Q_{out} = 105 ml/min	756	758	E = 5V I = 36 ma R_{out} = 6200 ohm-cm T = 18.5°C pH = 6.7 R_{in} = 4300 ohm-cm T_{in} = 18°C pH_{in} = 7.1 Q_{out} = 125 ml/min
4:00 PM	Voltage probe taken	757	759	Voltage probe taken
Apr. 3				
8:00 AM	E = 6.5V I = 36 ma R_{out} = 8000 ohm-cm T = 18.6°C pH = 7 R_{in} = 4300 ohm-cm T_{in} = 19.5°C pH_{in} = 7.4	773	775	E = 5V I = 36 ma R_{out} = 6600 ohm-cm T = 19°C pH = 7 R_{in} = 4300 ohm-cm T_{in} = 20°C pH_{in} = 7
9:00 AM	Voltage probe taken	774	776	Voltage probe taken
11:00 AM	Shut down system	776	778	Shut down system
Apr. 4	Augmenting sewage now with: PO_4; NO_3; ABS			
10:00 AM	Started system E = 4V I = 43 ma R_{out} = 3900 ohm-cm T = 16°C pH = 7 R_{in} = 3200 ohm-cm T_{in} = 15°C pH_{in} = 7.12 Q_{out} = 105 ml/min	776	778	Started system E = 4.2V I = 46 ma R_{out} = 3900 ohm-cm T = 16.5°C pH = 9.05 R_{in} = 3000 ohm-cm T_{in} = 17.5°C pH_{in} = 9.4 Q_{out} = 125 ml/min

TABLE A-2 (continued)

SYSTEMS OPERATION: EXHAUSTION RUNS - SERIES 2

Date/Time	System I	Hours of Operation		System II
Apr. 4 cont'd				
4:30 PM	E = 4.0V I = 33 ma R_{out} = 5000 ohm-cm T = 15°C pH = 7.1 R_{in} = 3300 ohm-cm T_{in} = 15°C pH_{in} = 7.2 Q_{out} = 105 ml/min	782.5	784.5	E = 4.3V I = 39 ma R_{out} = 4900 ohm-cm T = 15°C pH = 9 R_{in} = 3300 ohm-cm T_{in} = 16°C pH_{in} = 7.1 Q_{out} = 125 ml/min
4:45 PM	E = 6.5V I = 44 ma			E = 5.2V I = 48 ma
4:50 PM	Added 10 ml. conc. HCl to concentrate res.			Added 10 ml. conc. HCl to concentrate res.
4:55 PM	E = 6.5V I = 48 ma	783	785	E = 5V I = 50 ma
Apr. 5				
10:00 AM	E = 6.5V I = 38 ma R_{out} = 6000 ohm-cm T = 17.5°C pH = 7 R_{in} = 3600 ohm-cm T_{in} = 18°C pH_{in} = 7 Q_{out} = 105 ml/min	800	802	E = 4.5V I = 40 ma R_{out} = 4800 ohm-cm T = 18°C pH = 6.95 R_{in} = 3800 ohm-cm T_{in} = 18.5°C pH_{in} = 7 Q_{out} = 125 ml/min
Apr. 6				
6:30 AM	Voltage probe taken	820.5	822.5	Voltage probe taken
Noon	Shut down system	826	828	Shut down system
Apr. 8				
10:00 AM	Started system E = 7V I = 43 ma R_{out} = 8600 ohm-cm T = 16°C pH = 8 R_{in} = 4000 ohm-cm T_{in} = 17.5°C pH_{in} = 7.3 Q_{out} = 90 ml/min	826	828	Started system E = 5.2V I = 52 ma R_{out} = 5600 ohm-cm T = 16°C pH = 7.1 R_{in} = 3100 ohm-cm T_{in} = 20.5°C pH_{in} = 6.9 Q_{out} = 122 ml/min
11:45 AM	Voltage probe taken	828	830	Voltage probe taken
Apr. 9				
10:30 AM	Replaced pump	850.5		
4:20 PM	E = 7.5V I = 34 ma R_{out} = 6800 ohm-cm T = 16°C pH = 7.5 R_{in} = 4200 ohm-cm T_{in} = 18°C pH_{in} = 7.25 Q_{out} = 115 ml/min	856.5	858.5	E = 5.5V I = 38 ma R_{out} = 6200 ohm-cm T = 16°C pH = 7.4 R_{in} = 4700 ohm-cm T_{in} = 18°C pH_{in} = 7.2 Q_{out} = 123 ml/min
4:45 PM	Voltage probe taken	857	859	Voltage probe taken

TABLE A-2 (continued)

SYSTEMS OPERATION: EXHAUSTION RUNS - SERIES 2

Date/Time	System I	Hours of Operation		System II
Apr. 10				
10:00 AM	Shut down system	874	876	Shut down system
Apr. 11				
1:45 PM	Started system Added 5 ml conc. HCl to 6 L. conc. reservoir	874	876	Started system Added 5 ml conc. HCl to 6 L. conc. reservoir
3:45 PM	E = 7V I = 41 ma R_{out} = 5200 ohm-cm T_{out} = 18.5°C pH_{out} = 7.4 R_{in} = 4000 ohm-cm T_{in} = 15°C pH_{in} = 7.2 Q_{out} = 115 ml/min	876	878	E = 5V I = 40 ma R_{out} = 5200 ohm-cm T_{out} = 19.2°C pH_{out} = 8.9 R_{in} = 4000 ohm-cm T_{in} = 19°C pH_{in} = 9.2 Q_{out} = 125 ml/min
4:20 PM	Voltage probe taken	876.5	878.5	Voltage probe taken
Apr. 12				
8:00 AM	E = 7.82V I = 36 ma R_{in} = 3900 ohm-cm T_{in} = 17.2°C pH_{in} = 6.9 R_{out} = 5600 ohm-cm T_{out} = 16.1°C pH_{out} = 6.75 Q_{out} = 111 ml/min	892	894	E = 5.8V I = 42 ma R_{in} = 3380 ohm-cm T_{in} = 18°C pH_{in} = 6.8 R_{out} = 4650 ohm-cm T_{out} = 17.2°C pH_{out} = 6.8 Q_{out} = 125 ml/min
3:00 PM	E = 7.5V I = 38 ma R_{in} = 3400 ohm-cm T_{in} = 18.7°C pH_{in} = 6.9 R_{out} = 5400 ohm-cm T_{out} = 16°C pH_{out} = 6.8 Q_{out} = 150 ml/min	899	901	E = 5V I = 40 ma R_{in} = 3800 ohm-cm T_{in} = 19°C pH_{in} = 6.9 R_{out} = 5000 ohm-cm T_{out} = 15.6°C pH_{out} = 6.8 Q_{out} = 125 ml/min
3:15 PM	Changed Q_{out} to 130 ml/min			
3:30 PM	Voltage probe taken			Voltage probe taken
Apr. 13				
Noon	Shut down system	920	922	Shut down system
Apr. 18				
9:00 AM	Started system Polarization run	920	922	Started system Polarization run

TABLE A-2 (continued)

SYSTEMS OPERATION: EXHAUSTION RUNS - SERIES 2

Date/Time	System I	Hours of Operation		System II
Apr. 19 6:00 AM	E = 8.6V I = 50 ma R_{in} = 3100 ohm-cm T_{in} = 16°C pH_{in} = 6.8 R_{out} = 5100 ohm-cm T_{out} = 16°C pH_{out} = 6.9 Q_{out} = 130 ml/min	941	943	E = 4.8V I = 50 ma R_{in} = 3100 ohm-cm T_{in} = 18°C pH_{in} = 6.6 R_{out} = 4600 ohm-cm T_{out} = 18°C pH_{out} = 6.6 Q_{out} = 125 ml/min
Noon	Voltage probe taken	947	949	Voltage probe taken
12:30 PM	Added 10 ml conc. HCl to concentrate reservoir			
4:30 PM	E = 5V I = 34 ma R_{in} = 2900 ohm-cm T_{in} = 18.4°C pH_{in} = 7.05 R_{out} = 3800 ohm-cm T_{out} = 16.9°C pH_{out} = 6.85 Q_{out} = 129 ml/min			E = 4.5V I = 36 ma R_{in} = 3100 ohm-cm T_{in} = 20.2°C pH_{in} = 6.85 R_{out} = 3500 ohm-cm T_{out} = 19.6°C pH_{out} = 6.8 Q_{out} = 118 ml/min
5:00 PM	Shut down system	952	954	Shut down system
Apr. 22 10:40 AM	Started system E = 6.6V I = 44 ma R_{in} = 3600 ohm-cm T_{in} = 18.5°C pH_{in} = 7.2 R_{out} = 5400 ohm-cm T_{out} = 18°C pH_{out} = 7.1 Q_{out} = 130 ml/min	952	954	Started system E = 4.4V I = 42 ma R_{in} = 3300 ohm-cm T_{in} = 19.5°C pH_{in} = 7.2 R_{out} = 3800 ohm-cm T_{out} = 19°C pH_{out} = 6.75 Q_{out} = 123 ml/min
Apr. 23 7:05 AM	E = 7.5V I = 36 ma R_{in} = 3580 ohm-cm T_{in} = 17°C pH_{in} = 7.4 R_{out} = 4800 ohm-cm T_{out} = 14.4°C pH_{out} = 6.9 Q_{out} = 132.5 ml/min	972	974	E = 5.5V I = 36 ma R_{in} = 3600 ohm-cm T_{in} = 15.4°C pH_{in} = 7.4 R_{out} = 4760 ohm-cm T_{out} = 14°C pH_{out} = 7.4 Q_{out} = 112.5 ml/min
3:00 PM	E = 7V I = 36 ma R_{in} = 3600 ohm-cm T_{in} = 16°C pH_{in} = 6.5 R_{out} = 5000 ohm-cm T_{out} = 16°C pH_{out} = 6.8 Q_{out} = 125 ml/min	980	982	E = 5.3V I = 47 ma R_{in} = 3900 ohm-cm T_{in} = 16.5°C pH_{in} = 7.1 R_{out} = 4700 ohm-cm T_{out} = 17.5°C pH_{out} = 7.1 Q_{out} = 125 ml/min
3:30 PM	Voltage probe taken			Voltage probe taken

TABLE A-2 (continued)

SYSTEMS OPERATION: EXHAUSTION RUNS - SERIES 2

Date/Time	System I	Hours of Operation		System II
Apr. 24 8:00 AM	Shut down system	997	999	Shut down system
Apr. 25 11:00 AM	Dismantled system Inspected membranes			Dismantled system Inspected membranes
4:20 PM	Started system	997	999	Started system
4:45 PM	E = 4.8V I = 47 ma R_{out} = 5800 ohm-cm T_{out} = 17°C pH_{out} = 6.3 R_{in} = 3900 ohm-cm T_{in} = 15°C pH_{in} = 6.4 Voltage probe taken			E = 4.2V I = 62 ma R_{out} = 6000 ohm-cm T_{out} = 17.5°C pH_{out} = 9 R_{in} = 3900 ohm-cm T_{in} = 18°C pH_{in} = 9.5 Voltage probe taken
Apr. 26 8:00 AM	E = 4.8V I = 42.5 ma R_{out} = 4900 ohm-cm T_{out} = 17.9°C pH_{out} = 6.7 Q_{out} = 128 ml/min	1012.5	1014.5	E = 4.2V I = 45.5 ma R_{out} = 3700 ohm-cm T_{out} = 17.9°C pH_{out} = 7.15 Q_{out} = 110 ml/min
4:00 PM	E = 5.7V I = 45 ma R_{in} = 3400 ohm-cm T_{in} = 16.5°C pH_{in} = 6.4 R_{out} = 5800 ohm-cm T_{out} = 16.5°C pH_{out} = 6.45 Q_{out} = 125 ml/min Voltage probe taken	1020.5	1022.5	E = 4.7V I = 60 ma R_{in} = 3600 ohm-cm T_{in} = 18.5°C pH_{in} = 6.7 R_{out} = 4600 ohm-cm T_{out} = 18.5°C pH_{out} = 7 Q_{out} = 113 ml/min Voltage probe taken
Apr. 27 10:00 AM	E = 6V I = 46 ma R_{in} = 3800 ohm-cm T_{in} = 19°C pH_{in} = 7 R_{out} = 5300 ohm-cm T_{out} = 18.5°C pH_{out} = 6.6 Q_{out} = 125 ml/min	1038.5	1040.5	E = 4.4V I = 46 ma R_{in} = 3200 ohm-cm T_{in} = 19.7°C pH_{in} = 6.8 R_{out} = 4300 ohm-cm T_{out} = 18.5°C pH_{out} = 6.6 Q_{out} = 115 ml/min
10:30 AM	Shut down system	1039	1041	Shut down system
Apr. 29 9:30 AM	Started system	1039	1041	Started system

TABLE A-2 (continued)

SYSTEMS OPERATION: EXHAUSTION RUNS - SERIES 2

Date/Time	System I	Hours of Operation		System II
Apr. 29 cont'd				
2:00 PM	E = 5.7V I = 46 ma R_{in} = 4000 ohm-cm T_{in} = 20°C pH_{in} = 6.7 R_{out} = 6600 ohm-cm T_{out} = 19°C pH_{out} = 6.5 Q_{out} = 125 ml/min Voltage probe taken 10 ml. conc. HCl added to concentrate reservoir	1043.5	1045.5	E = 4.7V I = 69 ma R_{in} = 3500 ohm-cm T_{in} = 22°C pH_{in} = 6.9 R_{out} = 5400 ohm-cm T_{out} = 21°C pH_{out} = 6.8 Q_{out} = 110 ml/min Voltage probe taken 10 ml. conc. HCl added to concentrate reservoir
Apr. 30				
10:00 AM	E = 6.4V I = 45 ma R_{in} = 3900 ohm-cm T_{in} = 18°C pH_{in} = 6.85 R_{out} = 6600 ohm-cm T_{out} = 19.5°C pH_{out} = 6.4 Q_{out} = 128 ml/min	1063.5	1065.5	E = 4.8V I = 68 ma R_{in} = 4000 ohm-cm T_{in} = 18.5°C pH_{in} = 7.2 R_{out} = 6600 ohm-cm T_{out} = 18.7°C pH_{out} = 6.5 Q_{out} = 112 ml/min
May 1				
8:00 AM	Shut down system	1085.5	1087.5	Shut down system
May 2				
11:15 AM	Started system	1085.5	1087.5	Started system
1:00 PM	E = 5V I = 42 ma R_{in} = 2900 ohm-cm T_{in} = 16°C pH_{in} = 6.9 R_{out} = 4700 ohm-cm T_{out} = 17°C pH_{out} = 6.75 Q_{out} = 125 ml/min			E = 4V I = 52 ma R_{in} = 3100 ohm-cm T_{in} = 17°C pH_{in} = 8.5 R_{out} = 4400 ohm-cm T_{out} = 17.5°C pH_{out} = 6.35 Q_{out} = 115 ml/min
1:30 PM	Voltage probe taken			Voltage probe taken
May 3				
7:45 AM	E = 5V I = 43 ma R_{in} = 3100 ohm-cm T_{in} = 16.8°C pH_{in} = 7 R_{out} = 4100 ohm-cm T_{out} = 19.3°C pH_{out} = 6.75 Q_{out} = 125 ml/min	1109	1111	E = 4.2V I = 59 ma R_{in} = 3100 ohm-cm T_{in} = 19.8°C pH_{in} = 7.18 R_{out}= 3200 ohm-cm T_{out} = 18.1°C pH_{out} = 6.95 Q_{out} = 102 ml/min
Noon				Cathode stream plugged; stack disassembled and $CaCO_3$ buildup removed from cathode compartment

TABLE A-2 (continued)

SYSTEMS OPERATION: EXHAUSTION RUNS - SERIES 2

Date/Time	System I	Hours of Operation		System II
May 3 cont'd				
4:00 PM	Shut down system	1117	1119	Shut down system
May 6				
10:00 AM	Started system	1117	1119	Started system
11:30 AM	E = 4.5V I = 44 ma R_{in} = 2300 ohm-cm T_{in} = 19.5°C pH_{in} = 7.3 R_{out} = 3900 ohm-cm T_{out} = 18.5°C pH_{out} = 7.2 Q_{out} = 120 ml/min Conc Q_{out} = 102.5 ml/min Cath Q_{out} = 32.5 ml/min An. Q_{out} = 25 ml/min			E = 3.6V I = 56 ma R_{in} = 2400 ohm-cm T_{in} = 20°C pH_{in} = 7.2 R_{out} = 5600 ohm-cm T_{out} = 18.5°C pH_{out} = 5.0 Q_{out} = 122.5 ml/min Conc Q_{out} = 117 ml/min Cath Q_{out} = 27 ml/min An. Q_{out} = 27 ml/min
May 7				
8:00 AM	Shut down system	1139	1141	Shut down system Installed cartridge filter before charcoal column
May 9				
9:00 AM	Started system	1139	1141	Started system
	E = 3.4V I = 38 ma R_{in} = 2300 ohm-cm T_{in} = 21°C pH_{in} = 7 R_{out} = 3200 ohm-cm T_{out} = 21°C pH_{out} = 7 Q_{out} = 127 ml/min Voltage probe taken			E = 3.5V I = 50 ma R_{in} = 2300 ohm-cm T_{in} = 23°C pH_{in} = 6.7 R_{out} = 3500 ohm-cm T_{out} = 23°C pH_{out} = 6.7 Q_{out} = 130 ml/min Voltage probe taken Conc. reservoir: R = 170 ohm-cm pH = 2.9 System scaling
Noon				Conc. stream not being diluted. Cause of $CaCO_3$ scaling.
2:30 PM				Beginning 24 hour conc. recycle of 4000 ml. res.

TABLE A-2 (continued)

SYSTEMS OPERATION: EXHAUSTION RUNS - SERIES 2

Date/Time	System I	Hours of Operation		System II
May 10				
8:00 AM	Dilute stream not flowing. Feed line to dilute pump filled with sediment and slime growth. Line cleaned out and flow set at 130 ml/min.	1162	1164	
8:30 AM	E = 3.3V I = 39 ma R_{in} = 2800 ohm-cm T_{in} = 21°C pH_{in} = 6.5 R_{out} = 3200 ohm-cm T_{out} = 20°C pH_{out} = 6.5 Q_{out} = 133 ml/min			
2:30 PM			1170	Conc. volume 4415 ml. ΔV = 415 ml/day pH = 6.7 R = 260 ohm-cm T = 21°C
3:30 PM	E = 3.9V I = 42 ma R_{in} = 2400 ohm-cm T_{in} = 22°C pH_{in} = 7.25 R_{out} = 3600 ohm-cm T_{out} = 18°C pH_{out} = 7.2 Voltage probe taken	1169	1171	E = 3.4V I = 43 ma R_{in} = 3100 ohm-cm T_{in} = 19.5°C pH_{in} = 6.6 R_{out} = 3900 ohm-cm T_{out} = 19°C pH_{out} = 6.6 Voltage probe taken
May 11				
8:00 AM	E = 4.2V I = 34 ma R_{in} = 3950 ohm-cm T_{in} = 14°C pH_{in} = 7.15 R_{out} = 4600 ohm-cm T_{out} = 15°C pH_{out} = 7 Q_{out} = 130 ml/min	1186	1188	E = 3.9V I = 40 ma R_{in} = 3950 ohm-cm T_{in} = 15.5°C pH_{in} = 6.8 R_{out} = 4600 ohm-cm T_{out} = 15°C pH_{out} = 6.7 Q_{out} = 130 ml/min
10:00 AM	Voltage probe taken	1188	1190	Voltage probe taken
10:20 AM	Shut down			Shut down
May 13				
10:30 AM	Started system	1188.5	1190.5	Started system
3:00 PM	E = 5V I = 40 ma R_{in} = 2600 ohm-cm T_{in} = 20.5°C pH_{in} = 6.98 R_{out} = 3400 ohm-cm T_{out} = 20.5°C pH_{out} = 6.92			E = 5V I = 38 ma R_{in} = 2800 ohm-cm T_{in} = 22°C pH_{in} = 8.2 R_{out} = 3900 ohm-cm T_{out} = 21°C pH_{out} = 8.2

TABLE A-2 (continued)

SYSTEMS OPERATION: EXHAUSTION RUNS - SERIES 2

Date/Time	System I	Hours of Operation		System II
May 13 cont'd				
3:40 PM			1195	Added 8 ml. conc. HCl to 4.0L tap-water in conc. res. R = 142 ohm-cm pH = 1.95 T = 21°C
May 14				
3:40 PM			1219	Conc. ΔV = 300 ml/day R = 340 ohm-cm T = 21°C pH = 6.6
4:00 PM	E = 5.2V I = 38 ma R_{in} = 2700 ohm-cm T_{in} = 19°C pH_{in} = 7.1 R_{out} = 4000 ohm-cm T_{out} = 19.5°C pH_{out} = 7.1 Q_{out} = 125 ml/min	1218	1219	E = 4.2V I = 31 ma R_{in} = 2700 ohm-cm T_{in} = 20.5°C pH_{in} = 7.2 R_{out} = 3000 ohm-cm T_{out} = 20°C pH_{out} = 7
4:15 PM	Voltage probe taken			Voltage probe taken
4:30 PM	Shut down	1218.5	1219.5	Shut down
May 16				
4:30 PM	Started system	1218.5	1219.5	Started system
	Added 5 drops conc. H_2SO_4 to 4.0L conc. res. pH = 3.2 T = 18°C R = 2800 ohm-cm			Added 5 drops conc. H_2SO_4 to 4.0L conc. res. pH = 3.7 T = 17.5°C R = 3300 ohm-cm
May 17				
9:00 AM	E = 5.25V I = 38 ma R_{in} = 3250 ohm-cm T_{in} = 18.9°C pH_{in} = 6.9 R_{out} = 4300 ohm-cm T_{out} = 18.3°C pH_{out} = 6.9 Q_{out} = 130 ml/min	1235	1236	E = 4.2V I = 54 ma R_{in} = 3500 ohm-cm T_{in} = 18.8°C pH_{in} = 7 R_{out} = 5000 ohm-cm T_{out} = 18.6°C pH_{out} = 6.8 Q_{out} = 135 ml/min
3:30 PM	E = 5.2V I = 43 ma R_{in} = 3000 ohm-cm T_{in} = 23°C pH_{in} = 6.75 R_{out} = 4000 ohm-cm T_{out} = 23.5°C pH_{out} = 6.8 Q_{out} = 130 ml/min	1241.5	1242.5	E = 4.2V I = 45 ma R_{in} = 3300 ohm-cm T_{in} = 23°C pH_{in} = 6.7 R_{out} = 3900 ohm-cm T_{out} = 23.5°C pH_{out} = 6.7
4:00 PM	Voltage probe taken	1242	1243	Voltage probe taken
4:30 PM	Conc. ΔV = 225 ml/day R = 410 ohm-cm T = 24°C pH = 7.5 Shut down	1242.5	1243.5	Conc. ΔV = 275 ml/day R = 333 ohm-cm T = 24°C pH = 7.4 Shut down

TABLE A-2 (continued)

SYSTEMS OPERATION: EXHAUSTION RUNS - SERIES 2

Date/Time	System I	Hours of Operation		System II
May 20				
11:00 AM	Started system	1242.5	1243.5	Started system
1:00 PM	E = 5V I = 47 ma R_{in} = 2800 ohm-cm T_{in} = 23.7°C pH_{in} = 7.15 R_{out} = 3700 ohm-cm T_{out} = 24.7°C pH_{out} = 7.05 Q_{out} = 125 ml/min	1244.5	1245.5	E = 5.2V I = 45 ma R_{in} 3300 ohm-cm T_{in} = 14.7°C pH_{in} = 7.1 R_{out} = 3700 ohm-cm T_{out} = 24.3°C pH_{out} = 9.1 Q_{out} = 132 ml/min
3:00 PM	Added 5 drops conc. H_2SO_4 to 4.0L tap-water in conc. reservoir R = 2800 ohm-cm T = 24°C pH = 3.7	1246.5	1247.5	Added 5 drops conc. H_2SO_4 to 4.0L tap-water in conc. reservoir R = 2800 ohm-cm T = 24.4°C pH = 3.5
May 21				
3:00 PM	Conc. ΔV = 225 ml/day R = 310 ohm-cm T = 23°C pH = 7.6	1270.5	1271.5	Conc. ΔV = 230 ml/day R = 320 ohm-cm T = 22.5°C pH = 7.6
3:30 PM	E = 4.8V I = 40 ma R_{in} = 3290 ohm-cm T_{in} = 21.5°C pH_{in} = 7.3 R_{out} = 4000 ohm-cm T_{out} = 21.5°C pH_{out} = 7.25 Q_{out} = 125 ml/min Voltage probe taken	1271	1272	E = 5.3V I = 40 ma R_{in} = 3000 ohm-cm T_{in} = 22°C pH_{in} = 7.3 R_{out} = 3600 ohm-cm T_{out} = 22°C pH_{out} = 7.3 Q_{out} = 130 ml/min Voltage probe taken
5:00 PM	Conc. stream pH = 5.5 R = 3000 ohm-cm T = 21°C	1272.5	1273.5	Conc. stream pH = 5.15 R = 3000 ohm-cm T = 22°C
May 22				
8:00 AM	Conc. pH = 7.2 Added 2.0 ml conc. H_2SO_4 pH = 4.2	1287.5	1288.5	Conc. pH = 7.3 Added 25 drops conc. H_2SO_4 pH = 4.5
10:00 AM	Voltage probe taken	1289.5	1290.5	Voltage probe taken
10:30 AM	E = 5.3V I = 35 ma R_{in} = 4400 ohm-cm T_{in} = 20.7°C pH_{in} = 7.3 R_{out} = 5200 ohm-cm T_{out} = 21°C pH_{out} = 7.1 Q_{out} = 125 ml/min	1290	1291	E = 5V I = 38 ma R_{in} = 3800 ohm-cm T_{in} = 21.5°C pH_{in} = 7.1 R_{out} = 4500 ohm-cm T_{out} = 21°C pH_{out} = 7.1 Q_{out} = 132 ml/min

TABLE A-2 (continued)

SYSTEMS OPERATION: EXHAUSTION RUNS - SERIES 2

Date/Time	System I	Hours of Operation		System II
May 22 cont'd 1:00 PM	Conc. ΔV = 180 ml/day R = 380 ohm-cm T = 23°C pH = 6.3 Shut down	1292.5	1293.5	Conc. ΔV = 175 ml/day R = 700 ohm-cm T = 24°C pH = 5.7 Shut down
May 23 8:30 AM	Started system	1292.5	1293.5	Started system
9:00 AM	Conc. res. 4.0L tap water plus 3 drops conc. H_2SO_4 R = 3300 ohm-cm T = 21°C pH = 5.5	1293	1294	Conc. res. 4.0L tap-water plus 3 drops conc. H_2SO_4 R = 3200 ohm-cm T = 21°C pH = 5.5
10:30 AM	Conc. reservoirs drained to prepare for Nitrogen survey run in both systems			
3:00 PM	Returned to normal operation - both systems			
May 24 9:00 AM	E = 6V I = 34 ma Rin = 3000 ohm-cm Tin = 18°C pH_{in} = 7.0 Rout = 4400 ohm-cm Tout = 16°C pH_{out} = 7.0 Qout = 125 ml/min	1317	1318	E = 6V I = 50 ma Rin = 3000 ohm-cm Tin = 18°C pH_{in} = 7.0 Rout = 4000 ohm-cm Tout = 17°C pH_{out} = 7.05 Qout = 125 ml/min
May 25 9:00 AM	E = 6.9V I = 38 ma Rin = 3100 ohm-cm Tin = 20.5°C pH_{in} = 7.2 Rout = 4100 ohm-cm Tout = 19.5°C pH_{out} = 7.15 Qout = 125 ml/min	1341	1342	E = 6.2V I = 53 ma Rin = 3100 ohm-cm Tin = 23°C pH_{in} = 7.05 Rout = 4600 ohm-cm Tout = 18.5°C pH_{out} = 7.1 Qout = 125 ml/min
10:05 AM	Voltage probe taken	1342	1343	Voltage probe taken
10:15 AM	Shut down			Shut down
May 27 11:00 AM	Started system	1342	1343	Started system
3:00 PM	Added H_2SO_4 to conc. res. pH = 5.4	1346	1347	Added H_2SO_4 to conc. res. pH = 5.4
3:30 PM	E = 6.4V I = 43 ma Rin = 2900 ohm-cm Tin = 20.5°C pH_{in} = 7.2 Rout = 3200 ohm-cm Tout = 21°C pH_{out} = 7.2 Qout = 125 ml/min			E = 5.2V I = 50 ma Rin = 2600 ohm-cm Tin = 21°C pH_{in} = 6.8 Rout = 3500 ohm-cm Tout = 22°C pH_{out} = 6.8 Qout = 125 ml/min
May 28	Shut down systems to clean out dilute feed lines and honeycomb filters. (Slime deposits).			

TABLE A-2 (continued)

SYSTEMS OPERATION: EXHAUSTION RUNS - SERIES 2

Date/Time	System I	Hours of Operation		System II
May 28 cont'd				
10:00 AM	Started system	1363	1364	Started system
5:00 PM	E = 6.3V I = 48 ma R_{in} = 2400 ohm-cm T_{in} = 22.5°C pH_{in} = 7.3 R_{out} = 3000 ohm-cm T_{out} = 23°C pH_{out} = 7.35 Q_{out} = 125 ml/min	1370	1371	E = 5.8V I = 51 ma R_{in} = 2350 ohm-cm T_{in} = 23°C pH_{in} = 7.5 R_{out} = 3000 ohm-cm T_{out} = 24°C pH_{out} = 7.5 Q_{out} = 125 ml/min
5:15 PM	Voltage probe taken			Voltage probe taken
May 29				
8:00 AM	E = 6.3V I = 41 ma R_{in} = 2250 ohm-cm T_{in} = 19.7°C pH_{in} = 7.2 R_{out} = 3500 ohm-cm T_{out} = 19°C pH_{out} = 7.2 Q_{out} = 125 ml/min	1385	1386	E = 5.7V I = 46 ma R_{in} = 2500 ohm-cm T_{in} = 19.5°C pH_{in} = 7.2 R_{out} = 3000 ohm-cm T_{out} = 19°C pH_{out} = 7.2 Q_{out} = 122 ml/min
9:00 AM	Conc. ΔV = 525 ml/day R = 300 ohm-cm T = 19.5°C pH = 5.8	1386	1387	Conc. ΔV = 500 ml/day R = 280 ohm-cm T = 19.4°C pH = 6.3
3:30 PM	Shut down	1392.5	1393.5	Shut down
May 30				
Noon	Cathode compartment plugged with what appears to be $CaSO_4$ deposition		1393.5	Started system
1:00 PM			1394.5	E = 5V I = 52 ma
May 31				
8:00 AM			1413.5	E = 5.6V I = 37 ma R_{in} = 2500 ohm-cm T_{in} = 24°C pH_{in} = 7.35 R_{out} = 3100 ohm-cm T_{out} = 21.8°C pH_{out} = 7.4 Q_{out} = 127 ml/min
4:00 PM	Disassembled stack and cleaned out cathode compartment			
4:30 PM	Started system E = 2.3V I = 66 ma	1392.5		

TABLE A-2 (continued)

SYSTEMS OPERATION: EXHAUSTION RUNS - SERIES 2

Date/Time	System I	Hours of Operation		System II
June 1				
1:00 PM	E = 2.5V I = 55 ma Shut down	1413	1442.5	E = 5.5V I = 59 ma Shut down
June 3				
8:00 AM	Started system	1413	1442.5	Started system
10:00 AM	E = 3.3V I = 39 ma R_{in} = 2050 ohm-cm T_{in} = 28°C pH_{in} = 7.2 R_{out} = 2750 ohm-cm T_{out} = 28°C pH_{out} = 6.5 Q_{out} = 125 ml/min	1415	1444.5	E = 6.5V I = 45 ma R_{in} = 2000 ohm-cm T_{in} = 28°C pH_{in} = 7.6 R_{out} = 2500 ohm-cm T_{out} = 28°C pH_{out} = 8 Q_{out} = 125 ml/min
June 4				
8:00 AM	Shut down	1437	1466.5	Shut down
June 10				
8:00 AM	Started system	1437	1466.5	Started system
June 12				
10:00 AM	E = 5.2V I = 41 ma R_{in} = 3200 ohm-cm T_{in} = 16°C pH_{in} = 7.4 R_{out} = 3400 ohm-cm T_{out} = 16.5°C pH_{out} = 7.25 Q_{out} = 130 ml/min	1487	1516.5	E = 6V I = 29 ma R_{in} = 3300 ohm-cm T_{in} = 16°C pH_{in} = 7.4 R_{out} = 3500 ohm-cm T_{out} = 16.8°C pH_{out} = 7.25 Q_{out} = 125 ml/min
10:30 AM	Shut down	1487.5	1517	Shut down
June 20	Disassembled stacks from both systems			
	Small amounts of $CaCO_3$ and $CaSO_4$ scaling in cathode compartment. Small amounts of slime in dilute compartments. Brown slime in conc. compartments.			$CaCO_3$ and $CaSO_4$ build-up in cathode compartment. Next adjacent conc. compartment had heavy scaling. Next two conc. compartments away from cathode had very light scaling. Negligible slime deposits in diluting compartments

TABLE A-3.

PRODUCT/WASTE RATIO STUDIES - SYSTEMS OPERATION

DATE/ TIME	SYSTEM	DILUTE			CONC.			ANODE			CATHODE			MAKE-UP		ACID RES.			
		Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	pH	Conc ml	Cath ml	E V	I ma
June 21																			
12 Noon	I II				Initial equipment test - both systems		3.5 8.3						9.2 10.2					10 10	200 200
PM	I				Polarization study														
5:00 PM	I II				Shut down both systems														
June 24																			
8:15 AM	I II				Started both systems														
3:00 PM	I II	352 372			525 420 Acid feed 3.6N H_2SO_4 - both systems		5.7 1.9	90 105			110 125		2.9 5.4						
3:35 PM	I II													56 43					
3:40 PM	I II													32 23					
3:45 PM	I II				Shut off conc. acid feed - both systems		1.4 1.4												

TABLE A-3 (continued)

PRODUCT/WASTE RATIO STUDIES - SYSTEMS OPERATION

DATE/ TIME	SYSTEM	DILUTE			CONC.			ANODE			CATHODE			MAKE-UP		ACID RES.			
		Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	pH	Conc ml	Cath ml	E V	I ma
June 24 cont'd																			
3:50 PM	I II		4.8 2.5			5.0 2.8			5.9 5.0	4.6 3.3		6.0 6.5	2.2 2.0						
4:00 PM	I II	455 457	5.0 2.8	5.7 5.9	525 190	5.4 2.5	1.5 1.8	60 75	2.5 2.0	3.9 3.1	75 75	1.9 3.5	7.6 9.0	35 33	7.2				
5:00 PM	I II				Started cathode acid feeds - both systems													10.5 11	200 380
5:10 PM	I II				Stopped cathode acid feeds - both systems								1.6 1.6						
5:30 PM	I II			6.1 6.2			1.8 2.1						6.5 9.5		7.3 7.3				
6:00 PM	I II												8.3 10.0					11 11	180 330
June 25																			
7:45 AM	I				System I piping broken by vandalism during night. Tanks drained. System I shut down.								8.3						
7:45 AM	II												8.3						

TABLE A-3 (continued)

PRODUCT/WASTE RATIO STUDIES - SYSTEMS OPERATION

DATE/ TIME	SYSTEM	DILUTE			CONC.			ANODE			CATHODE			MAKE-UP		ACID RES.			
		Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	pH	Conc ml	Cath ml	E V	I ma
June 25 cont'd																			
9:00 AM	II	550	4.8	5.9	285 Conc. overflow recycled to tank	6.2	8.3	74	2.5	2.8	70	3.4	1.1	26	6.2			10.8	120
9:45 AM	II				Acid feed pump on		8.2												
10:00 AM	II				(0.36M H_2SO_4)		1.8						1.7						
10:20 AM	II												1.85					10	360
10:25 AM	II						1.85											11	400
10:37 AM	II						1.9						2.3						
10:44 AM	II						1.9						2.5						
11:00 AM	II						1.9						2.8						
11:05 AM	II	550	5.8	3.9	435	5.8	1.9	65	2.6	2.6	75	3.2	3.2	36	5.8			11.2	420
11:30 AM	II				Voltage probe taken														
11:55 AM	II				Cathode acid feed increased		2.0						9.4					11.2	480
12:15 PM	II				Conc. overflow vol. = 4.0L								1.3						

TABLE A-3 (continued)

PRODUCT/WASTE RATIO STUDIES - SYSTEMS OPERATION

DATE/ TIME	SYSTEM	DILUTE			CONC.			ANODE			CATHODE			MAKE-UP		ACID RES.			
		Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	pH psi	Conc ml	Cath ml	E V	I ma
June 25 cont'd																			
12:40 PM	II				Overflow rate = 33 ml/min		2.2						2.1						
1:00 PM	II						2.2						2.3					11.2	450
1:10 PM	II						2.3						2.6						
1:20 PM	II		8.2																400
1:45 PM	II	395	8.6																
2:15 PM	II	375			Shut down - removed small bits of charcoal and scale from stack														
2:30 PM	II				Started system														
3:30 PM	II	510		4.3			2.8						9.5						
3:40 PM	II						2.8						3.5						
4:20 PM	II													36				11	440
5:00 PM	II						2.3			2.8			1.7					11	490
5:30 PM	II	550	6.0	3.6	625	5.5	2.4						2.1			1950	1900	11.8	530
June 26																			
8:00 AM	II				Added conc. overflow (14L) to dilute influent tank											1900	1000	11.8	280

TABLE A-3 (continued)

PRODUCT/WASTE RATIO STUDIES - SYSTEMS OPERATION

DATE/ TIME	SYSTEM	DILUTE			CONC.			ANODE			CATHODE			MAKE-UP		ACID RES.			
		Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	pH	Conc ml	Cath ml	E V	I ma
June 26 cont'd																			
8:30 AM	II	380	8.2	3.2	490	3.8	2.7	80	3.0	2.9	60	3.0	2.7	46	3.3				
9:00 AM	II																	11.8	360
4:30 PM	II	315	8.0	3.3	570	5.5	2.9	75	2.8	3.0	80	3.7	2.9	39	3.4				
4:55 PM	II				Conc. overflow vol. = 8L											1900	700	12	330
5:00 PM	II				Shut down														
June 27																			
10:00 AM	I II				Started both systems														
12 Noon	I II			6.2 5.8											6.9			7.5 8	130 175
12:10 PM	I II			6.5 6.4														5 5	90 90
12:30 PM	I II	390 350	3.6 3.8	6.2 6.8	375 500	3.0 4.0	2.4 4.0	175 70	4.0 3.0	5.8 5.4	80 84	1.0 4.0	1.7 3.3	33 35	6.8 6.9				
4:15 PM	I II	380 420	3.5 4.3	6.2 7.0	425 440	3.6 3.6	3.8 2.5	82 155	4.5 5.4	6.1 6.1	86 102	2.7 5.8	7.4 8.4	39 42	6.8 7.6				
5:00 PM	I II				Conc. overflow vols. 5.8L 4.0L														

TABLE A-3 (continued)

PRODUCT/WASTE RATIO STUDIES - SYSTEMS OPERATION

DATE/ TIME	SYSTEM	DILUTE			CONC.			ANODE			CATHODE			MAKE-UP		ACID RES.			
		Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	pH	Conc ml	Cath ml	E V	I ma
June 27 cont'd																			
5:30 PM	I II															1900 1750	1990 1750		
June 28																			
8:00 AM	I II															1690 1690	1690 1500		
8:30 AM	I II				Conc. overflow vols.			27L 12L											
9:00 AM	II	410	4.7	7.2	425 Increased cathode acid feed. Decreased make-up flow.	6.1	7.1	79	3.6	7.0	86	4.2	7.5	46	7.3				
10:00 AM	I II				Conc. overflow vols.			28L 13L											
10:05 AM	I II				Added both conc. overflow to dilute influent tanks. Waste overflow rate = 33 ml/min														
10:10 AM	II				Voltage probe indicates scaling													5.5	30
10:30 AM	II				Increased conc. acid feed. Added 0.2 ml conc. HCl to recycle stream		6.4						6.7						
10:45 AM	II			6.9											7.0			10	90

TABLE A-3 (continued)

PRODUCT/WASTE RATIO STUDIES - SYSTEMS OPERATION

DATE/ TIME	SYSTEM	DILUTE			CONC.			ANODE			CATHODE			MAKE-UP		ACID RES.			
		Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	pH	Conc. ml	Cath. ml	E V	I ma
June 28	cont'd																		
11:15 AM	I II	325	2.8	5.2 6.7	370	3.0	4.8	58	2.5	3.4	60	2.1	2.6	45	5.7 6.7			5.2 10	100 78
12 Noon	I II				Changeover from H_2SO_4 (0.36M) to HCl (0.36M) as acid feed														
12:10 PM	I II															2000 2000	2000 2000		
12:50 PM	II				Increased acid feed													10	70
1:00 PM	II															1900	1900		
1:05 PM	II																	10	195
1:07 PM	II																	10	210
1:08 PM	II																	5	90
4:00 PM	I II	375 350	3.5 3.0	5.2 6.7			3.7 5.7						2.7 2.2					5.5 5.5	95 90
5:00 PM	I II				Polarization Run														
5:30 PM	II	330	3.0	6.3	465	4.2	2.3	68	3.4	6.0	105	6.1	2.5	39	6.8			5.5	100

TABLE A-3 (continued)

PRODUCT/WASTE RATIO STUDIES - SYSTEMS OPERATION

DATE/ TIME	SYSTEM	DILUTE			CONC.			ANODE			CATHODE			MAKE-UP		ACID RES.			
		Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	pH	Conc ml	Cath ml	E V	I ma
June 28 cont'd																			
6:15 PM	I	450	4.5	5.6	510 Shut down System I. Conc. overflow vol. = 17L	5.0	5.3	84	5.2	3.6	90	3.0	2.7	45	5.6			5.0	80
6:30 PM	I II				Shut down System II. Conc. overflow vol. = 15.5L											1050 1600	1750 1300		
July 1																			
7:30 AM	I II				Started both systems. Added conc. overflow volumes to tanks														
8:45 AM	I II						2.9 2.1						2.8 2.1					5.0 5.2	95 130
10:00 AM	I II				Increased acid feed to System I		4.8 2.0						7.8 2.1						
10:15 AM	I												2.5						
10:30 AM	I	402	4.0	6.6	450	4.1	2.3	80	4.8	5.7	86	3.0	2.5	39	6.6			5.2	100
11:00 AM	II	355	2.4	6.4	365 Voltage probe taken	2.8	1.9	61	2.5	5.3	83	5.0	2.2	43	6.8			5.2	110
3:00 PM	I II															450 800	1570 700		
4:00 PM	I II				Polarization run														

TABLE A-3 (continued)

PRODUCT/WASTE RATIO STUDIES - SYSTEMS OPERATION

DATE/ TIME	SYSTEM	DILUTE			CONC.			ANODE			CATHODE			MAKE-UP		ACID RES.			
		Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	pH	Conc ml	Cath ml	E V	I ma
July 1 cont'd																			
4:30 PM	I II				Increased conc. acid feed - System I		6.3 2.3						1.8 2.5						
4:40 PM	I II						4.5									400 600	1500 600		
4:50 PM	I II				0.38M HCl added to acid res. - both systems											2000 2000	2000 2000		
5:00 PM	I II				Conc overflow volumes I = 15.7L II = 18.0L						Added conc. overflow vols. to tanks								
5:15 PM	I II						2.5 2.2						1.75 2.3					5.3 5.2	102 118
July 2																			
7:30 AM	I II				Conc. overflow volumes I = 20L II = 20L														
7:50 AM	I II				Added conc. overflow vols. to tanks											200 480	700 1800		
8:15 AM	I II			6.3 8.0			2.9 2.3						2.3 4.4		6.6 8.3				

TABLE A-3 (continued)

PRODUCT/WASTE RATIO STUDIES - SYSTEMS OPERATION

DATE/ TIME	SYSTEM	DILUTE			CONC.			ANODE			CATHODE			MAKE-UP		ACID RES.			
		Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	pH	Conc. ml	Cath ml	E V	I ma
July 2 cont'd																			
9:00 AM	I II	365 330	3.4 3.9	7.3 6.4	470 440 System I - R_{in} = 2000 ohm-cm; R_{out} = 2500 ohm-cm System II - R_{in} = 1970 ohm-cm; R_{out} = 2600 ohm-cm	4.0 3.6	2.2 2.2	70 68	3.6 5.0	6.5 3.3	86 80	5.0 3.5	4.3 2.5	38 38	7.6 6.4			5 5	120 95
9:30 AM	I															2000			
11:45 AM	I II															1820 200	490 1710		
12 Noon	I II				Added (0.38M) HCl - both systems											2000 2000	2000 2000		
1:50 PM	I II															1790 1790			
4:55 PM	I II				Conc. overflow vols. I = 18L II = 15L						Added conc. overflows to tanks								
5:15 PM	I II						2.4 3.1						3.0 2.5					5.2 5.0	115 140
July 3																			
7:30 AM	I II				Conc. overflow vols. I = 15L II = 21L													5.2 5.0	90 107

TABLE A-3 (continued)

PRODUCT/WASTE RATIO STUDIES - SYSTEMS OPERATION

DATE/ TIME	SYSTEM	DILUTE			CONC.			ANODE			CATHODE			MAKE-UP		ACID RES.			
		Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	pH	Conc ml	Cath ml	E V	I ma
July 3 cont'd																			
8:30 AM	I II				Increased acid feed to System I		6.0 3.5						8.0 1.9			1650 1510	1800 700		
9:00 AM	II				Cathode Acid Feed 0.944 ml/min Conc. Acid Feed 0.362 ml/min											1500	700		
10:00 AM	I II	340 275	6.3 3.5	5.7 7.1	340 350 System I - R_{in} = 1950 ohm-cm; R_{out} = 2600 ohm-cm System II - R_{in} = 1800 ohm-cm; R_{out} = 2400 ohm-cm	2.5 3.5	2.3 3.5	55 56	7.5 4.5	3.1 5.3	63 65	2.5 4.5	1.8 2.3	32.5 38	6.5 7.8			5 5	110 110
11:00 AM	I II				Conc. overflow rate I = 34 ml/min II = 27 ml/min						Voltage probe taken								
12:30 PM	I II				Conc. overflow vols; I = 20L II = 26L						Added conc. overflow vols. to tanks								
12:45 PM	I II			7.0 6.9	System I - R_{out} = 1500 ohm-cm System II - R_{out} = 1700 ohm-cm														
3:55 PM	I II				Both systems operating on recycle basis. Both conc. overflow and dilute effluent recycled to tanks.														
4:00 PM	I II															500 1300	1300 300		
4:10 PM	I II				Acid reservoirs 0.38N HCl - both systems											2000 2000	2000 2000	5 5.2	150 160

TABLE A-3 (continued)

PRODUCT/WASTE RATIO STUDIES - SYSTEMS OPERATION

DATE/ TIME	SYSTEM	DILUTE			CONC.			ANODE			CATHODE			MAKE-UP		ACID RES.			
		Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	pH	Conc. ml	Cath ml	E V	I ma
July 4																			
11:00 AM	I II				System I acid reservoirs filled to 2000 ml.											100 1100	100 830	5 5	80 35
12 Noon	I II			6.0 7.1	2.4 2.7 Increased cathode acid feed in System II					5.5 6.5			2.5 7.0		6.7 8.6				
12:05 PM	I II												2.8			1080	780	5 5	95 75
12:10 PM	I	390			Increased dilute flow rate. Conc. overflow = 28 ml/min									30					
12:10 PM	II	455			Conc. overflow = 31 ml/min									37		1850	2000		
July 5																			
12:30 PM	I II															190 780	300 150	5 5	97 85
12:35 PM	I II	440 440	10 6.6	6.0 7.0	3.0 3.0 System I - Overflow = 11 ml/min System II- Overflow = 26 ml/min										6.2 7.0	2000 1600	2000 2000		
1:00 PM	II				Shut off conc. make-up in System II														

TABLE A-3 (continued)

PRODUCT/WASTE RATIO STUDIES - SYSTEMS OPERATION

DATE/ TIME	SYSTEM	DILUTE			CONC.			ANODE			CATHODE			MAKE-UP		ACID RES.			
		Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	pH	Conc ml	Cath ml	E V	I ma
July 6																			
9:00 AM	I II						2.0 1.7						6.8 6.2			1400 525	2000 100		
9:30 AM	I II															2000 2000	2000 2000	5.2 5.2	95 97
10:00 AM	I	415	9.6	5.3	600	8.2	2.2	76	11.4	3.9	92	7.2	7.3	9.5	6.6				
11:00 AM	I II	 460	 6.5	 6.0	 550 Voltage probe taken - System II	 7.5	 1.9	 75	 8.5	 5.9	 105	 6.7	 8.5	 0				5.8 5.8	105 110
11:15 AM	I II			5.2 6.5	System I - R_{in} = 2300 ohm-cm; R_{out} = 3000 ohm-cm System II - R_{in} = 2500 ohm-cm; R_{out} = 3300 ohm-cm										6.4 7.2				
11:20 AM	I				Increased acid feed to System I cathode stream														
11:35 AM	I II				R_{conc} = 90 ohm-cm - System I R_{conc} = 50 ohm-cm - System II											1800 1825	1825 1780		
12 Noon	I II				Conc overflows: I = 14 ml/min II = 1.4 ml/min Took 1 liter sample from System II conc. reservoir														
12:30 PM	I II															1700 1775	1710 1700		
5:30 PM	I II				Pumping conc. overflow to tanks													5 5	100 100

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TABLE A-3 (continued)

PRODUCT/WASTE RATIO STUDIES - SYSTEMS OPERATION

DATE/ TIME	SYSTEM	DILUTE			CONC.			ANODE			CATHODE			MAKE-UP		ACID RES.			
		Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	pH	Conc ml	Cath ml	E V	I ma
July 6 cont'd																			
6:00 PM	I II				New load of sewage													5 5	65 38
July 7																			
3:00 PM	I II				System I conc. reservoir down 3 liters											1575 200	875 1700	5 5	60 115
3:10 PM	I II				System I conc. pump leaking. Shut down System I.											2000	1800		
July 8																			
8:30 AM	I II				System I conc. drained											1560 800	580 150		
8:35 AM	I II				System II cath. filled											0 800	580 1600		
9:55 AM	I				Disassembled System I. No scaling in cathode stream. Slight scale in adjacent conc. stream.														
12 Noon	I II				System I reassembled Shut down System II											600	900	4.8	105
	II				Disassembled System II. No scaling. Small slime buildup in dilute streams														

TABLE A-3 (continued)

PRODUCT/WASTE RATIO STUDIES - SYSTEMS OPERATION

DATE/ TIME	SYSTEM	DILUTE			CONC.			ANODE			CATHODE			MAKE-UP		ACID RES.			
		Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	pH	Conc ml	Cath ml	E V	I ma
July 8 cont'd																			
1:30 PM	I															0	580		
	II				Started System II											600	875		
1:40 PM	I				Started System I											2000	2000		
	II															2000	2000		
4:15 PM	I															1900	1900		
	II	480	4.6	6.4	550	5.0	1.8	100	0.5	6.2	120	6.6	6.0	0	6.9	1700	1850	5	95
					System II - R_{in} = 2500; R_{out}= 3350; R_{conc} = 50 ohm-cm.														
4:40 PM	I															1810	1880		
	II															1700	1810		
4:45 PM	I															2000	2000		
	II				System II conc. acid now 0.19N HCl											2000	2000		
5:00 PM	I	530	8.0	7.1	450	4.0	3.9	105	0.5	6.7	100	3.3	2.8	20	7.1			5	90
					System I - R_{in} = 2400; R_{out} = 2700; R_{conc} = 700 ohm-cm														
5:00 PM	I				Conc. overflow rate = 20 ml/min. Decreased dilute Q to 460 ml/min. Increased make-up to 46 ml/min.														
5:40 PM	I																	4.8	90
	II						2.0											5.0	95
					Voltage probes taken														

TABLE A-3 (continued)

PRODUCT/WASTE RATIO STUDIES - SYSTEMS OPERATION

DATE/ TIME	SYSTEM	DILUTE			CONC.			ANODE			CATHODE			MAKE-UP		ACID RES.			
		Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	pH	Conc ml	Cath ml	E V	I ma
July 9																			
8:00 AM	I						7.0									2000			60
					Increased conc. acid feed. Increased conc. recycle flow rate														
9:00 AM	I						3.2						2.6			1900	800		
	II						2.4						2.5			675	400		
9:30 AM	I				Recycled Anode and Cathode streams to tanks after adding NaOH													5	105
	II																	5	90
10:30 AM	I															1790	725		
	II	415	5.0	7.7	730	7.9	2.3	60	0	5.9	62.5	3.5	2.4	0	7.7	525	200	5	105
					System II - R_{in} = 2200; R_{out} = 2500; R_{conc} = 68 ohm-cm														
10:35 AM	I				Acid now 0.19N HCl											2000	2000		
	II															2000	2000		
11:00 AM	I	385	7.7	11.1	600	6.6	3.0	50	0	10.5	50	0.5	11.0	38	11.1			5	110
					System I - R_{in} = 1000; R_{out} = 1200; R_{conc} = 520 ohm-cm. Conc. overflow = 36 ml/min. System II - Conc. acid feed = 1.36 ml/min. Conc overflow = 2.22 ml/min														
3:30 PM	I			7.2			3.0						3.1		7.4	1600	1700		
	II			8.7			2.4						3.4		8.7	1550	1500		
					System I - R_{in} = 1200; R_{out} = 1400 ohm-cm System II- R_{in} = 1000; R_{out} = 1200 ohm-cm														
4:00 PM	I				Acid 0.19N HCl											2000	2000		
	II															2000	2000		

TABLE A-3 (continued)

PRODUCT/WASTE RATIO STUDIES - SYSTEMS OPERATION

DATE/ TIME	SYSTEM	DILUTE			CONC.			ANODE			CATHODE			MAKE-UP		ACID RES.			
		Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	pH	Conc ml	Cath ml	E V	I ma
July 9 cont'd																			
4:15 PM	I II				Voltage probe taken													4.9 5.0	200 230
July 10																			
8:45 AM	I II															800 600	1790 420		
9:00 AM	I	480	7.0	4.0	480 System I - R_{in} = 900; R_{out} = 1200; R_{conc} = 135 ohm-cm	5.3	2.5	65	0	2.9	70	2.5	2.8	36.4	4.0			5	250
9:00 AM	II	365	4.4	6.5	490 System II - R_{in} = 1350; R_{out} = 1680; R_{conc} = 500 ohm-cm	4.5	2.3	98	0.5	3.5	79	4.6	3.2	0	6.7			5	220
					System II: Cath., Dil., conc. flow increased. System I: Make-up increased. Added NaOH to tank (I)														
10:30 AM	I II				Polarization runs											2000 2000	2000 2000	10 10	
12 Noon	I																	7.5	
12:15 PM	I			6.0	System I - R_{in} = 900; R_{out} = 1300 ohm-cm								7.0		6.3			7.5	400
12:30 PM	I			4.5	System I - R_{in} = 900; R_{out} = 900; R_{conc} = 160 ohm-cm		2.5						3.2		5.6			5	220
12:45 PM	I				Voltage probe taken													4.8	250

TABLE A-3 (continued)

PRODUCT/WASTE RATIO STUDIES - SYSTEMS OPERATION

DATE/ TIME	SYSTEM	DILUTE			CONC.			ANODE			CATHODE			MAKE-UP		ACID RES.			
		Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	pH	Conc ml	Cath ml	E V	I ma
July 10 cont'd																			
4:00 PM	I II				New load of sewage											1520 1400	1590 1600		
4:05 PM	I II			7.0 6.4	Voltage probe - System II		2.6 2.0			6.5 3.6			5.7 5.1		7.0 6.8	2000 2000	2000 2000	5 4.9	120 120
July 11																			
9:45 AM	I II				Voltage probes taken														120 110
10:10 AM	I II						2.5 2.0						2.8 5.2			890 500	550 1025		
11:00 AM	II				Cathode acid feed increased to 0.38N HCl								1.3						
12:45 PM	I II				Concs I & II and Cath I Acid = 0.19N HCl Cath. II Acid = 0.38N HCl											2000 2000	2000 2000		
2:00 PM	I II			6.8 6.8	System I - R_{in} = 1200; R_{out} = 1500; R_{conc} = 170 ohm-cm System II - R_{in} = 1600; R_{out} = 1820; R_{conc} = 41 ohm-cm		2.5 2.1			6.5 6.0			5.9 2.6		7.6 6.8			5 5	130 190
3:45 PM	I II						2.6 2.2						4.6 2.4			1675 1680	1700 1400		
4:00 PM	I	500			Changed cathode acid feed to 0.38N HCl. Conc overflow = 50 ml/min.														

TABLE A-3 (continued)

PRODUCT/WASTE RATIO STUDIES - SYSTEMS OPERATION

DATE/ TIME	SYSTEM	DILUTE			CONC.			ANODE			CATHODE			MAKE-UP		ACID RES.			
		Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	pH	Conc ml	Cath ml	E V	I ma
July 11 cont'd																			
4:15 PM	I II				Cath. 0.38N HCl Conc. 0.19N HCl											2000 2000	2000 2000		
July 12																			
8:00 AM	II				Filled to 2175 ml												200		
8:30 AM	I II				System I - R_{conc} = 130 ohm-cm; T = 15°C System II - R_{conc} = 39 ohm-cm; T = 18.1°C		2.6 2.2						8.5 2.1						
9:00 AM	I II	415 485		7.0 7.0	Overflow rate = 42 ml/min								2.2	4.6	7.3 7.1		1500	5 5	160 260
					System I - R_{out} = 1490 ohm-cm; T = 14.9°C: R_{in} = 1280 ohm-cm; T = 13.8°C System II - R_{out} = 1490 ohm-cm; T = 18.3°C: R_{in} = 1280 ohm-cm; T = 16.0°C														
9:30 AM	I II															1190 610	1300 1990		
11:50 AM	I II						2.7 2.4						1.6 2.7			1000 400	840 1775		
12:15 PM	I II															2000 2000	2000 2000		

TABLE A-3 (continued)

PRODUCT/WASTE RATIO STUDIES - SYSTEMS OPERATION

DATE/ TIME	SYSTEM	DILUTE			CONC.			ANODE			CATHODE			MAKE-UP		ACID RES.			
		Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	pH	Conc ml	Cath ml	E V	I ma
July 12 cont'd																			
1:00 AM	I	480	7.1	6.8	580	6.5	3.2	60		4.0	94	4.0	2.3	41	6.8			4.4	160
	II	480	7.2	6.9	520	6.7	2.5	110	0.75	5.2	60	3.5	2.4	0	7.0			4.8	220
					System I - Conc. overflow = 42 ml/min														
2:00 PM	I II				Voltage probes taken													4.4 4.8	160 240
6:00 PM	I II															1620 1650	1450 1590		
6:15 PM	I II				Conc. = 0.19N HCl Cath. = 0.38N HCl											2175 2175	2175 2175		
July 13																			
12 Noon	I II				System I conc. pump leaking Shut down both systems													5.5 5.5	125 130
July 15																			
8:00 AM	I II				Started both systems														
9:45 AM	I II				Conc. = 0.19N HCl Cath. = 0.38N HCl											2175 2175	2175 2175	5 5	120 160
10:15 AM	II				System II - R_{in} = 1790; R_{out} = 2250; R_{conc} = 60 ohm-cm						60	3.6	2.1						
	II	550	8.0	7.6	500	6.0	2.4	5		6.2	90	5.5	2.6	0	7.7				

TABLE A-3 (continued)

PRODUCT/WASTE RATIO STUDIES - SYSTEMS OPERATION

DATE/ TIME	SYSTEM	DILUTE			CONC.			ANODE			CATHODE			MAKE-UP		ACID RES.			
		Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	pH	Conc ml	Cath ml	E V	I ma
July 15 cont'd																			
10:30 AM	I I	470	8.5	8.7	560 System I - Cathode pH controlled by varying flow rate	7.0	2.4	80	1.0	6.9	90 58	3.4 1.0	9.5 1.8	45	8.0				
	I I										78	2.4 2.0	8.7 3.3						
11:45 AM	I II				Voltage probes taken													4.9 5.4	130 170
11:50 AM	I II				System I conc. overflow rate = 46 ml/min											2000 1900	2025 1925		
4:00 PM	I II						2.8 2.3						2.6 2.7			2175 2175	2175 2175		
July 16																			
8:30 AM	I II	412 400		7.0 7.3	Sample Y taken		2.6 2.4						1.8 2.9	37 0	7.6 7.6			5 5	110 150
10:30 AM	I II				Filled all acid res. to 2175 ml.											700 575	375 550		
2:30 PM	I II												2.6 3.0						

TABLE A-3 (continued)

PRODUCT/WASTE RATIO STUDIES - SYSTEMS OPERATION

DATE/ TIME	SYSTEM	DILUTE			CONC.			ANODE			CATHODE			MAKE-UP		ACID RES.			
		Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	pH	Conc ml	Cath ml	E V	I ma
July 16 cont'd																			
3:15 PM	I II															1810 1720	1700 1800		
3:30 PM	I I				Voltage probes taken													4.6 5.4	120 160
3:50 PM	I II	400 450	10 10	7.1 7.0	560 640 System I - conc. overflow = 40 ml/min	7.6 9.5	3.4 2.4	70 66	2.3	6.3 5.9	100 70	5 4.2	2.9 3.0	40 0	7.1 7.3				
4:30 PM	I I															2175 2175	2175 2175		
5:00 PM	I				R_{out} = 1750 ohm-cm; T_{out} = 28.5°C: R_{in} = 1500 ohm-cm; T_{in} = 27.2°C: R_{conc} = 390 ohm-cm; T_{conc} = 28.3°C.														
	II				R_{out} = 1900 ohm-cm; T_{out} = 29.0°C: R_{in} = 1600 ohm-cm; T_{in} = 28.0°C: R_{conc} = 49 ohm-cm; T_{conc} = 29.0°C.														
July 17																			
9:30 AM	I II	390 480	6.5 6.4				2.5 2.0				100 70		2.2 2.4	44 0	7.0 6.8	900 780	1150 1300	5 5	120 195
2:45 PM	I II						3.0 2.0						2.2 2.7			2175 2175	2175 2175		
3:15 PM	I II				Voltage probes taken													5 5	140 200

TABLE A-3 (continued)

PRODUCT/WASTE RATIO STUDIES - SYSTEMS OPERATION

DATE/ TIME	SYSTEM	DILUTE			CONC.			ANODE			CATHODE			MAKE-UP		ACID RES.			
		Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	pH	Conc ml	Cath ml	E V	I ma
July 17 cont'd																			
3:45 PM	I II	420 460	10.6 11.6	6.4 6.3	500 500	6.5 7.5	2.9 2.0	70 80	2.5	3.6 3.5	86 90	3.2 5.0	2.3 6.3	44 0	6.8 6.7				
					System I - R_{in} = 1350; R_{out} = 1720 ohm-cm System II- R_{in} = 1400; R_{out} = 1680 ohm-cm														
	II										73	4.0	3.3						
4:30 PM	I II				Conc. 0.19N HCl Cath. 0.38N HCl											2175 2175	2175 2175		
July 18																			
9:30 AM	I II						2.9 1.8				96 72		2.0 2.3			1000 690	1600 1100	5 5	100 160
					New load of sewage														
10:30 AM	II	460	11.0	6.7	480	6.5	1.8	82		6.2	90	4.8	2.5	0	7.4			5	160
					T_{in} = 29.4; T_{out} = 29.7; T_{conc} = 30.5°C R_{in} = 1700; R_{out} = 1900; R_{conc} = 42 ohm-cm														
10:45 AM	II			6.7											7.4			0	0
11:00 AM	II				Added NaOH to conc. stream														
11:20 AM	II			6.8			2.7												0
					Added more NaOH														

TABLE A-3 (continued)

PRODUCT/WASTE RATIO STUDIES - SYSTEMS OPERATION

DATE/ TIME	SYSTEM	DILUTE			CONC.			ANODE			CATHODE			MAKE-UP		ACID RES.			
		Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	ph	Q ml/ min	P psi	pH	Q ml/ min	pH	Conc ml	Cath ml	E V	I ma
July 18 cont'd																			
11:30 AM	II			7.3	Added HCl to conc. stream		7.0												0
11:35 AM	II			7.0	Back to normal operation		3.1												
12:30 PM	II				R_{conc} = 44 ohm-cm; T_{conc} = 31.5°C														
12:50 PM	I	400	10.6	6.7	530	6.7	1.7	80	4.0	6.4	94	4.3	2.1	40	6.8			5	100
1:15 PM	I II				Voltage probes taken													5.4 5.4	100 140
4:10 PM	I II															2175 2175	2175 2175		
July 19																			
8:00 AM	II	500		6.7			1.7							0	7.3			8	240
8:30 AM	I	400		5.8			1.8							44	6.5			6.6	120
9:00 AM	I II										40 75		1.4 2.1			1100 600	400 1180		
11:00 AM	I II															2175 2175	2175 2175		

TABLE A-3 (continued)

PRODUCT/WASTE RATIO STUDIES - SYSTEMS OPERATION

DATE/ TIME	SYSTEM	DILUTE			CONC.			ANODE			CATHODE			MAKE-UP		ACID RES.			
		Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	pH	Conc ml	Cath ml	E V	I ma
July 19 cont'd																			
1:00 PM	II	460	13.0	7.0	600	10.5	2.6	90		5.9	70	4.9	1.8	0	7.4			8.0	220
1:30 PM	I	350	9.0	5.5	500	6.5	2.0	70	9.3	4.0	75	2.8	3.5	44	6.4			6.6	120
2:30 PM	I II				Voltage probes taken													6.7 7.7	120 250
4:00 PM	I II				Cath. acid = 0.76N HCl Conc. acid = 0.19N HCl											2175 2175	2175 2175		
July 20																			
1:30 PM	I II				Filled all reservoirs to 2175 ml.		2.0 1.7						7.5 1.9			400 200	2000 150	7.5 8.5	110 215
2:00 PM	I												1.9						
July 21																			
2:00 PM	I II				Filled all reservoirs to 2175 ml											300 700	150 2000	7.5 8.5	125 215
July 22																			
8:00 AM	I II						1.9 7.0						7.0 7.0			600 1500	200 200		

TABLE A-3 (continued)

PRODUCT/WASTE RATIO STUDIES - SYSTEMS OPERATION

DATE/ TIME	SYSTEM	DILUTE			CONC.			ANODE			CATHODE			MAKE-UP		ACID RES.			
		Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	pH	Conc ml	Cath ml	E V	I ma
July 22 cont'd																			
8:30 AM	I II															2175 2175	2175 2175	7.0 7.5	140 38
9:30 AM	I II												1.5 1.4						
10:30 AM	II	400	20	6.8	600 T_{in} = 23.8; T_{out} = 25.6; T_{conc} = 25.5°C R_{in} = 2100; R_{out} = 2300; R_{conc} = 55 ohm-cm.	18	4.0	62	3.3	4.3	70	5.0	2.1	0	6.9			7.5	170
10:45 AM	II	420	20⁺		Voltage probe taken													7.8	170
11:05 AM	II				Disassembled System II. Removed small amounts of charcoal and slime from system.														
1:10 PM	II	840	12		System II now functioning normally														
2:00 PM	II	720	9.6	6.9	600 T_{in} = 23.7; T_{out} = 25.0; T_{conc} = 26.2°C R_{in} = 2100. R_{out} = 2300; R_{conc} = 55 ohm-cm	8.4	5.3	60		3.2	62	3.5	2.0		6.9			7.8	220
2:30 PM	II				Voltage probe taken													7.8	200
3:00 PM	II				Polarization study														
4:00 PM	I	550	17⁺	4.9	500 Voltage probe taken	17⁺	2.3	52.	8.0	2.8	60	2.0	2.1	50	5.3			6.6	140

TABLE A-3 (continued)

PRODUCT/WASTE RATIO STUDIES - SYSTEMS OPERATION

DATE/ TIME	SYSTEM	DILUTE			CONC.			ANODE			CATHODE			MAKE-UP		ACID RES.			
		Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	pH	Conc ml	Cath ml	E V	I ma
July 22 cont'd																			
4:30 PM	I				Disassembled System I. Removed small amounts of slime from anode and dilute streams														
5:00 PM	I	720			System functioning normally. Polarization study														
6:00 PM	I II				Added NaOH to conc. reservoirs														
6:30 PM	I II															2175 2175	2175 2175		
July 23																			
8:00 AM	I II	640	20^+	7.1	Scaling in conc. (II). Acid had stopped.	20^+	7.0								7.2	1020 2000	800 2000	12	200
9:30 AM	I II	692 720		6.4 6.9	Conc. HCl added to conc. stream (II). Corrected scaling		2.0 2.1			2.5 2.0			1.6 1.4	60 0	6.8 7.2			8 12	320 420
10:20 AM	I II	680 680	10.5 14.5	6.2 6.9	720 600 Polarization study - System II	9.7 11.5	2.5 2.1	52 50		3.0 2.3	78 78	2.8 5.0	7.0 1.7	70 0	6.6 7.2			8 12	320 410

TABLE A-3 (continued)

PRODUCT/WASTE RATIO STUDIES - SYSTEMS OPERATION

DATE/ TIME	SYSTEM	DILUTE			CONC.			ANODE			CATHODE			MAKE-UP		ACID RES.			
		Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	pH	Conc ml	Cath ml	E V	I ma
July 23 cont'd																			
11:00 AM	I II				Voltage probes taken													7.8 12.0	320 405
1:00 PM	I II II				After decreasing acid feed		2.4 1.7						1.3 1.2 1.5			2175 2175	2175 2175	7.8 9.5	310 380
4:30 PM	I II						2.4 2.0						1.3 1.5			2175 2175	2175 2175		
July 24																			
8:30 AM	I II				Acid to cathode increased								2.0 11.0			780 1250	700 1700	9.5	300
8:50 AM	I II				System II - conc acid = 0.964 ml/min conc overflow = 2.67 ml/min		2.1 2.0												
10:00 AM	II	680	20$^+$	6.8	840	19.0	1.9	64		3.0	64	5.5	1.9	0	7.1			9.5	320
10:30 AM	I	840	18$^+$	5.9	1040 Voltage probes taken	18$^+$	2.2	25		2.8	52	1.5	1.9	60	6.2			7.8	180
11:15 AM	I II				Stopped acid feed to conc. streams in both systems to increase pH														
1:30 PM	I II						3.0 2.0						1.8 1.3						

TABLE A-3 (continued)

PRODUCT/WASTE RATIO STUDIES - SYSTEMS OPERATION

DATE/ TIME	SYSTEM	DILUTE			CONC.			ANODE			CATHODE			MAKE-UP		ACID RES.			
		Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	pH	Conc ml	Cath ml	E V	I ma
July 24 cont'd																			
3:00 PM	I II															2175 2175	2175 2175		
July 25																			
8:00 AM	I II				Cathode acid feed in System I stopped during night. Increased cathode acid feed in both systems.														
	I				Dilute stream flow rate decreased. Pressure drop increased during night.														
8:30 AM	I II																	8.2 9.5	160 260
9:00 AM	I II						2.1 2.3						1.2 1.4			600 1500	2000 1600		
9:15 AM	I II				Stopped acid feed to both conc. streams.														
10:45 AM	I II						2.3 2.3												
11:10 AM	I II						2.4 2.4												

TABLE A-3 (continued)

PRODUCT/WASTE RATIO STUDIES - SYSTEMS OPERATION

DATE/ TIME	SYSTEM	DILUTE			CONC.			ANODE			CATHODE			MAKE-UP		ACID RES.			
		Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	pH	Conc ml	Cath ml	E V	I ma
July 25 cont'd																			
11:30 AM	I	472		6.0											6.3				
	II	480		6.2											6.8				
					System I - Dilute R = 2100 ohm-cm; Dilute T = 35°C; Makeup R = 1900 ohm-cm; T = 33°C System II - Dilute R = 3000 ohm-cm; Dilute T = 33°C; Makeup R = 2200 ohm-cm; T = 33°C														
11:45 AM	I						2.5											8.2	140
	II						2.4											3.5	270
					Added NaOH to System II conc. stream														
12 Noon	I						2.6												
	II						3.9												
					Started System II acid feed to conc. stream. Acid = .05N HCl														
12:15 PM	I				Added NaOH to conc. stream														
12:30 PM	I						4.5												
					Started acid feed to conc. stream														
12:45 PM	I						4.1												
	II						3.4												
					System II - changed acid to 0.025N HCl														
2:00 PM	I						3.0												
	II						3.0												

TABLE A-3 (continued)

PRODUCT/WASTE RATIO STUDIES - SYSTEMS OPERATION

DATE/ TIME	SYSTEM	DILUTE			CONC.			ANODE			CATHODE			MAKE-UP		ACID RES.			
		Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	pH	Conc ml	Cath ml	E V	I ma
July 25 cont'd																			
3:00 PM	I II						2.7 3.3												
					Changed cartridge filter in both systems														
3:30 PM	I II				Voltage probes taken - both systems														
4:00 PM	I II				Filled all acid reservoirs											2175 2175	2175 2175		
4:30 PM	I II	720 720		6.0 6.5			2.7 3.6						1.2 1.4						
5:00 PM	I				Voltage probe taken														
July 26																			
8:00 AM	I				Dilute stream (tank 3) ran out during night												100	18	160
9:00 AM	I				Pumping from tank 4 to System I												2175	18.0	420
9:15 AM	I II						1.4 3.6									800 1700	2150 1100		
9:45 AM	I	720		4.7	1000		1.6			2.0			1.0	64	6.2				
					System I - R_D = 2400 ohm-cm; R_C = 100 ohm-cm; R_{MU} = 1600 ohm-cm; T_D = 31.8°C; T_C = 31.5°C; T_{MU} = 30°C														

TABLE A-3 (continued)

PRODUCT/WASTE RATIO STUDIES - SYSTEMS OPERATION

DATE/ TIME	SYSTEM	DILUTE			CONC.			ANODE			CATHODE			MAKE-UP		ACID RES.			
		Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	pH	Conc ml	Cath ml	E V	I ma
July 26	cont'd																		
10:05 AM	I				Shut off current														
10:15 AM	I			5.1														0	0
10:18 AM	I																	10	260
10:20 AM	II	680		6.8	1000 System II - R_D = 2350 ohm-cm; R_C = 29 ohm-cm; R_{MU} = 1700 ohm-cm; T_D = 36°C; T_C = 35°C; T_{MU} = 33.4°C		5.0	40		2.3	44		1.3		7.0			10.5	270
11:45 AM	I II						2.0 5.5												
Noon	I II												1.2 1.5						
4:45 PM	I II				System II - T_C = 38.7°C; R_C = 2700 ohm-cm Voltage probes taken - both systems		2.7 6.3												
5:20 PM	II	360			1000 Added HCl to conc. to pH 4		6.6											10.5	240
5:30 PM	II			7.1	R_D = 2350 ohm-cm; R_{MU} = 1900 ohm-cm, T_D = 37°C; T_{MU} = 34.5°C										7.1				
5:50 PM	II				After acid added to conc. stream													10.5	420
6:00 PM	II				Voltage probe taken		2.0						1.0						

TABLE A-3 (continued)

PRODUCT/WASTE RATIO STUDIES - SYSTEMS OPERATION

DATE/ TIME	SYSTEM	DILUTE			CONC.			ANODE			CATHODE			MAKE-UP		ACID RES.			
		Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	pH	Conc ml	Cath ml	E V	I ma
July 26 cont'd																			
6:30 PM	I II						2.8						1.4			2175 2175	2175 2175	6	300
July 27																			
7:30 AM	I II						3.5 2.0						2.2 1.3			1600 1700	400 600	7 12	140 320
8:00 AM	I II				Filled all acid reservoirs											2175 2175	2175 2175		
8:30 AM	II	680		6.45	R_D = 3600 ohm-cm; T_D = 33.5°C;		2.0	R_C = 23 ohm-cm; T_C = 33°C;			R_{MU} = 2500 ohm-cm; T_{MU} = 31°C				7.0				
8:45 AM	I	920		6.8	R_D = 2100 ohm-cm; T_D = 33°C; Voltage probes taken		4.0	R_C = 410 ohm-cm; T_C = 33.6°C;			R_{MU} = 1850 ohm-cm; T_{MU} = 31.5°C			80	6.8			7	140
3:15 PM	I II				Filled all acid reservoirs to 2175											1750 1950	18 1800		
4:45 PM	I II						3.9 2.9						2.0 1.9						
5:45 PM	I II				Filled all acid reservoirs to 2175 ml.													8 12.8	170 360

TABLE A-3 (continued)

PRODUCT/WASTE RATIO STUDIES - SYSTEMS OPERATION

DATE/ TIME	SYSTEM	DILUTE			CONC.			ANODE			CATHODE			MAKE-UP		ACID RES.			
		Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	P psi	pH	Q ml/ min	pH	Conc ml	Cath ml	E V	I ma
July 28																			
12:30 PM	I II			7.0 5.7			4.3 2.6						2.6 8.3		7.1 6.5	880 1610	180 1410	8 12.5	180 365
					Filled all acid reservoirs to 2175 ml.														
12:40 PM	II												2.4						
July 29																			
8:00 AM	I II						3.8 2.8									1200 300	900 180	8 11.5	190 320
					System I - R_C = 130 ohm-cm; T_C = 38°C System II - R_C = 22 ohm-cm; T_C = 39.3°C Stopped acid feed to conc. II stream														
9:15 AM	I	728		7.1			4.0			2.9			1.7	65	7.1			8	180
					System I - R_D = 1950 ohm-cm; R_C = 170 ohm-cm; R_{MU} = 1150 ohm-cm T_D = 36°C: T_C = 36.5°C; T_{MU} = 34.8°C														
9:15 AM	II	550		5.8			3.8			2.1			2.0	0	6.5			11.5	310
					System II - R_D = 3500 ohm-cm; R_C = 22 ohm-cm; R_{MU} = 2300 ohm-cm; T_D = 37.4°C; T_C = 38°C; T_{MU} = 34.5°C Started acid feed to conc. II stream.														
10:30 AM	I II				Voltage probes taken														
11:00 AM	I II				Shut down both systems - end of tests.														

APPENDIX B

Analytical Procedures And Data

TABLE B-1.

RESULTS OF ANALYSES - SYSTEM I, SERIES 1

Constituent*	12/30/62 Influent	12/30/62 Effluent	12/30/62 Influent	12/30/62 Effluent	1/4/63 Influent	1/4/63 Effluent
Sodium, Na^+	16.0	6.0		5.0	50.0	20.0
Potassium, K^+	5.0	2.0		2.5	8.0	4.5
Ammunium, NH_4^+	4.3	2.4				
Calcium, Ca^{+2}	9.7	3.9		2.4	9.7	17.4
Magnesium, Mg^{+2}	1.7	0.6		3.5	5.7	5.8
Chloride, Cl^-	20.8	12.5		15.0	45.6	30.3
Bicarbonate, HCO_3^-					117.1	86.6
Nitrate, NO_3^-	1.4	0.9		1.0	1.2	0.4
Sulfate, $SO_4^=$	32.0	16.5		90.0	24.4	24.0
Phosphate, $PO_4^{\equiv}$	4.4	4.0		23.0	21.3	15.5
ABS	0.22	0.24		1.07		
COD	25.2	21.1		0.95	257	107
Alkalinity, meq/l					1.92	1.42
pH	7.5	7.1		2.9	6.4	6.4
Resistivity, ohm-cm	3680	8100			1690	3300
Temperature, °C	11	11				
I, ma	50				115	
Q, ml/min	120				120	

*In mg/liter, except as otherwise indicated.

TABLE B-2.

RESULTS OF ANALYSES - SYSTEM I, SERIES 2

Constituent*	1/29/63		2/5/63		2/8/63		2/12/63		2/15/63	
	Influent	Effluent	Influent	Effluent	Influent	Effluent	Influent	Effluent	Influent	Effluent
Sodium, Na^+	40.0	11.0	21.2	10.8			29.0	12.3		
Potassium, K^+	5.3		10.5	3.5						
Ammonium, NH_4^+	15.0	2.5	19.4	negl.	25.5	10.6	20.6	10.3	33.3	27.0
Calcium, Ca^{+2}	10.0	2.6	11.6	3.9			9.9	7.0		
Magnesium, Mg^{+2}	3.5	3.0	1.2	1.1			1.4	0.2		
Chloride, Cl^-	24.6	5.8	27.2	8.7	66.3	62.3	36.9	8.5	49.0	26.5
Bicarbonate, HCO_3^-	86.6	9.8	108.6	28.7	213.5	151.3	104.9	42.7	186.7	192.8
Nitrate, NO_3^-	1.0	1.2	0.8	0.7			2.2	1.4		
Sulfate, $SO_4^=$	27.9	13.6	16.0	13.0			36.3	17.5		
Phosphate, $PO_4^≡$	6.0	3.0	8.7	7.0			6.0	2.5		
Total Ions	219.9		225.2	77.4						
ABS	2.65	2.2	1.7	1.43			0.96	1.02		
COD		51.2	31.9	84.0			118	460		
Alkalinity, meq/l	1.42	0.16	1.78	0.47	3.50	2.48	1.72	0.70	3.06	3.16
pH	7.2	5.7	7.4	6.3	7.6	7.3	7.05	6.5	7.48	7.12
Resistivity, ohm-cm	4190	15100	3600	8800	2500	3520	3600	8700	2750	4100
Temperature, °C	16.5	17.8	18.0	20.5	13.0	15.0	17.0	18.0	10.5	14.0
I, ma	100		80				76			
Q, ml/min	115		110				115			

TABLE B-2 (continued)

Constituent*	2/18/63		2/19/63		2/27/63		3/8/63		3/12/63	
	Influent	Effluent	Influent	Effluent	Influent	Effluent	Influent	Effluent	Influent	Effluent
Sodium, Na^+			21.4	23.8			38.0	29.4		
Potassium, K^+			3.0	2.0			7.0	4.0		
Ammonium, NH_4^+			15.1	negl.	26.75	18.85	20.5	15.3	16.0	12.0
Calcium, Ca^{+2}	13.7	6.3	12.5	7.5	14.7	11.3	11.5	8.8	13.2	10.1
Magnesium, Mg^{+2}	2.7	1.5	0.42	0.66	0.64	0.64	2.8	2.4	1.93	1.46
Chloride, Cl^-			26.2	14.3	23.1	16.8	37.0	23.8	31.6	21.8
Bicarbonate, HCO_3^-			135.4	96.4	142.7	128.1	137.9	106.8	107.4	96.6
Nitrate, NO_3^-			8.5	8.5	3.0	4.0	2.5	1.5	8.5	negl.
Sulfate, $SO_4^{=}$					13.0	11.0	19.3	14.6	22.4	19.2
Phosphate, $PO_4^{\equiv}$			8.5	8.0	11.5	9.5	20.0	18.0	7.0	7.0
Total Ions							296.5	224.6		
ABS			0.86	0.84	0.68	0.70	0.62	0.60	0.36	0.72
COD			79.3	56.6	41.0 44.0	29.0 41.0	148.0 119.0	141.0 106.0	19.7 44.7	20.2 44.4
Alkalinity, meq/l			2.22	1.58	2.34	2.10	2.26	1.75	1.76	1.52
pH			7.5	6.9	7.2	7.1	7.3	7.1	7.1	7.2
Resistivity, ohm-cm			3500	4800	2900	3700	3300	4400	3000	4800
Temperature, °C			18.5	19.0	17.0	14.5	14.0	13.0	18.5	18.5
Turbidity, SiO_2							50.0	40.0		
I, ma			60		24		30		20	
Q, ml/min			120		112.5		115		113	

TABLE B-2 (continued)

Constituent*	3/15/63 Influent	3/15/63 Effluent	3/19/63 Influent	3/19/63 Effluent	3/26/63 Influent	3/26/63 Effluent	3/29/63 Influent	3/29/63 Effluent	4/2/63 Influent	4/2/63 Effluent
Sodium, Na^+										
Potassium, K^+										
Ammonium, NH_4^+	11.7	11.2	10.33	7.68	18.0	10.9	5.3	3.7	6.4	4.0
Calcium, Ca^{+2}	14.6	8.5	13.9	9.4						
Magnesium, Mg^{+2}	2.73	2.88	1.7	1.25						
Hardness, Ca^{+2}					16.5	10.5	19.0	11.8	17.6	7.5
Chloride, Cl^-	38.6	26.2	30.4	20.3	27.9	14.0	36.6	18.5	29.6	12.5
Bicarbonate, HCO_3^-	96.6	86.6	97.6	78.7	63.4	50.6	61.0	40.3	57.3	33.6
Nitrate, NO_3^-	3.0	2.0	9.0	8.5	15.0	4.5	19.5	5.5	24.5	8.0
Sulfate, $SO_4^=$	23.0	21.0	21.5	18.0	21.0	18.0	17.0	18.0	24.0	19.4
Phosphate, $PO_4^{\equiv}$	7.0	6.7	2.0	2.0	6.0	2.0	8.0	6.0	3.5	1.5
ABS	1.18	0.95	0.12	0.18	0.37	0.49	0.48	0.60	0.1	0.2
COD	61.5 60.3	70.0 70.0	29.0 19.0	21.0 19.0			35.0 38.0	35.0 33.0	43.5 39.6	13.1 19.8
Alkalinity, meq/l	1.584	1.420	1.60	1.29	1.04	0.83	1.00	0.66	0.94	0.55
pH	7.1	6.9	7.4	7.4	6.83	6.7	6.7	6.5	6.85	6.65
Resistivity, ohm-cm	3300	4300	3000	4600	3600	7000	4000	7800	4000	8000
Temperature, °C	17.5	15.5	17.4	13.0	21.5	18.0	17.5	16.4	18.0	18.0
Turbidity, SiO_2			13.0	10.0			25.0	20.0		
I, ma	20		19		32		33		37	
Q, ml/min	105		105		110		105		105	

TABLE B-2 (continued)

Constituent *	4/5/63		4/9/63		4/12/63		4/19/63		4/23/63	
	Influent	Effluent	Influent	Effluent	Influent	Effluent	Influent	Effluent	Influent	Effluent
Sodium, Na^+										
Potassium, K^+										
Ammonium, NH_4^+	9.0	5.0			18.2	7.2	16.0	10.35	51.3	48.43
Hardness, Ca^{+2}	19.2	11.6	16.1	8.0	15.1	9.9	14.3	9.06	14.3	9.84
Chloride, Cl^-	35.0	16.4	26.9	13.25	33.2	18.1	29.2	14.3	30.0	18.3
Bicarbonate, HCO_3^-	93.9	62.2	64.7	34.2	85.4	56.1	101.9	69.5	103.1	81.7
Nitrate, NO_3^-	4.0	3.5	19.5	8.0	0.5	negl.	6.5	2.5	2.75	0.25
Sulfate, $SO_4^=$	36.0	21.0	21.0	17.2	24.0	24.0	42.0	29.4	22.4	24.0
Phosphate, $PO_4^{\equiv}$	18.5	13.0	23.5	20.0	39.0	29.0	31.0	27.5	23.5	23.0
ABS	0.12	0.08	0.34	0.31	4.20	3.36	2.8	2.4	2.92	2.60
COD	52.0 60.0	16.0 20.0	26.1 24.3	26.1 24.3	67.0 67.0	70.0 70.0	1694 1690	74.0 73.5	40.7 42.6	52.5 52.0
Alkalinity, meq/l	1.54	1.02	1.06	0.56	1.40	0.92	1.67	1.14	1.69	1.34
pH	7.0	7.0	7.25	7.5	6.9	6.75	7.05	6.85	7.4	6.9
Resistivity, ohm-cm	3600	6000	4200	6800	3900	5600	2900	3800	3500	4800
Temperature, °C	18.0	17.5	18.0	16.0	17.2	16.1	18.4	16.9	17.0	14.4
Turbidity, SiO_2					47.0	25.0	17.0	10.0	17.0	10.0
Organic N	4.1	2.1	4.7	2.6	1.1	2.2				
I, ma	38.0		34.0		36.0		34.0		36.0	
Q, ml/min	105.0		115.0		111.0		129.0		137.5	

TABLE B-2 (continued)

Constituent*	4/30/63		5/3/63		5/17/63		5/24/63		5/29/63		6/12/63	
	Influent	Effluent	Influent	Effluent	Influent	Effluent	Influent	Effluent	Influent	Effluent	Influent	Effluent
Sodium, Na^+					37.5	27.4	39.0	34.0	46.2	42.0		
Potassium, K^+					13.7	11.9	16.5	13.0	20.5	17.5		
Ammonium, NH_4^+	40.4	6.02	18.87	11.40	11.5	18.25	17.0	13.8	18.75	16.25	30.2	49.8
Hardness, Ca^{+2}	14.9	7.72	15.1	9.04	14.3	9.3	14.5	12.3	15.0	8.36	13.23	10.0
Chloride, Cl^-	22.2	9.45	29.2	21.4	24.2	15.0	27.3	19.0	28.6	17.2	27.1	23.3
Bicarbonate, HCO_3^-	98.8	63.4	152.5	109.8	98.2	87.8	117.1	100.0	150.1	130.5	126.3	114.7
Nitrate, NO_3^-	37.5	12.5	2.0	2.0	16.5	10.0	7.5	3.0	12.0	6.0		
Sulfate, $SO_4^=$	22.0	18.2	31.6	29.0	27.0	24.0	43.0	27.0	30.0	25.0		
Phosphate, $PO_4^{\equiv}$	24.0	21.0	31.0	30.0	42.5	36.0	33.5	28.5	29.5	27.5		
Total Ions					285.4	239.65	315.4	250.6	350.45	290.31		
ABS	3.68	3.48	3.52	1.76	3.5	1.8	10.0	8.0	7.76	7.65		
COD	35.7 30.4	28.5 28.5	81.5 81.5	81.5 79.5	76.8 80.0	80.0 82.6	70.0 76.0	70.0 71.5	79.5 79.5	77.5 79.5		
Alkalinity, meq/l	1.62	1.04	2.50	1.80	1.61	1.44	1.92	1.64	2.46	2.14	2.07	1.88
pH	6.85	6.4	7.0	6.78	6.9	6.9	7.0	7.0	7.2	7.2	7.4	7.25
Resistivity, ohm-cm	3900	6600	3100	4100	3250	4300	3000	4400	2550	3500	3200	3400
Temperature, °C	18.0	19.5	16.8	19.3	18.9	18.3	18.0	16.0	19.7	19.0	16.0	16.5
Turbidity, SiO_2	8.0	4.0	6.0	2.5	18.0	15.0	42.0	33.0	17.0	17.0		
Organic N							3.0	2.8	4.4	4.2		
I, ma	45.0		43.0		38.0		34.0		41.0		41	
Q, ml/min	128.0		125.0		130.0		125.0		125.0		130	

*In mg/liter, except as otherwise indicated.

TABLE B-3.

RESULTS OF ANALYSES - SYSTEM I, PRODUCT/WASTE RATIO STUDY

Constituent*	6/28/63 Influent	6/28/63 Effluent	6/28/63 Conc	7/2/63 Influent	7/2/63 Effluent	7/2/63 Conc	7/12/63 Influent	7/12/63 Effluent	7/12/63 Conc
Sodium, Na^+	50.0	45.0	350.0	38.0	23.0	168.0	86.0	74.0	650.0
Potassium, K^+	12.2	10.0	87.0	45.0	24.0	140.0	57.0	48.5	550.0
Ammonium, NH_4^+	35.0	9.8	57.0	23.9	15.2	77.0	24.6	18.6	191.0
Hardness, Ca^{+2}	14.7	10.0	68.0	16.0	10.8	82.0	8.2	7.2	257.2
Chloride, Cl^-	24.5	12.7	122.0	73.0	46.1	645.0	111.0	100.0	1860.0
Bicarbonate, HCO_3^-	15.9	9.2	0	59.8	34.8	0	105.0	85.4	0
Nitrate, NO_3^-	5.5	2.1	400.0	11.8	8.9	47.0	24.3	23.7	600.0
Sulfate, $SO_4^=$	135.0	115.0	850.0	41.0	37.4	75.0	66.0	58.0	56.8
Phosphate, $PO_4^{\equiv}$	33.0	29.0	170.0	28.0	22.5	52.5	31.0	31.0	72.0
Total Ions	325.8	242.8	2074.0	336.5	222.7	1286.5	513.1	446.4	4237.0
ABS				> 2.0	1.8	0.6	1.8	1.8	2.0
COD							45.8 48.0	45.5 47.5	27.8 27.8
Alkalinity, meq/l	0.26	0.15	0						
pH	5.5	5.2		6.6	6.3	2.9	7.3	7.0	2.6
Resistivity, ohm-cm	2500	3800		2000	2500		1280	1490	130
Temperature, °C	18.0	29.0					13.8	14.9	15
Turbidity, SiO_2	15.0	15.0		70.0	60.0	106	9.5	9.5	
Organic N							2.4	4.1	
TDS							420	340	4352
I, ma		100			120			160	
Q, ml/min		325			365			415	

TABLE B-3 (continued)

Constituent*	7/16/63			7/19/63			7/23/63			7/29/63		
	Influent	Effluent	Conc	Influent	Effluent	Conc	Influent	Effluent	Conc	Influent	Effluent	Conc
Sodium, Na^+	88.5	65.0	539.0	50.0	35.0	400.0	104.0	84.0	875.0	100.0	80.0	400.0
Potassium, K^+	33.4	26.6	275.0	18.5	12.5	220.0	35.5	27.0	330.0	15.0	12.0	250.0
Ammonium, NH_4^+	26.0	31.3	203.0	20.8	20.4	258.0	21.35	18.53	292.0	18.6	15.45	180.0
Hardness, Ca^{+2}	10.7	7.8	435.7	13.8	13.6	260.0	16.4	12.8	232.0	10.4	7.6	240.0
Chloride, Cl^-	65.1	56.8	3140.0	89.0	70.7	2200.0	38.4	31.8	2750.0	85.0	69.0	2919.0
Bicarbonate, HCO_3^-	152.0	124.0	0	64.1	16.5	0	114.0	94.4	0	62.22	54.9	0
Nitrate, NO_3^-	22.5	9.0	120.0	<2.5	<2.5	20.0	45.0	25.0	360.0	0	0	62.5
Sulfate, $SO_4^=$	28.2	26.4	57.0	39.0	38.0	118.0	45.0	44.0	350.0	32.0	30.0	775.0
Phosphate, $PO_4^{\equiv}$	30.0	30.0	72.0	23.0	22.0	88.0	45.0	43.0	92.0	25.0	20.0	58.0
Total Ions	456.4	376.9	4841.7	320.7	231.2	3564.0	464.65	380.53	5281	348.2	289.0	4884.5
ABS	3.2	3.5	2.2	1.8	2.0	1.8	1.4	1.6	1.8	2.2	2.0	1.8
COD	81.7 81.7	76.0 75.0		72.5 75.0	69.4 68.4		83.0 83.0	83.0 82.0		25.9 26.2	25.9 25.5	
Alkalinity, meq/l										1.02	0.9	
pH	7.6	7.0	2.6	6.5	5.8	1.8	6.8	6.4	2.0	7.1	7.1	4.0
Resistivity, ohm-cm	1600	1850	155	1900	2475	87	1300	1600	90	1150	1950	170
Temperature, °C	25.9	27.2	27.5	28.0	29.5	29.5	22.4	23.6	24.2	34.8	36.0	36.5
Turbidity, SiO_2	51.0	51.0	45	35.0	36.0	> 50	33.0	33.0	> 51	10.0	9.0	27.5
Organic N	3.4	2.9	28.0	3.3	2.2		3.5	3.2	29.9	3.6	3.5	
TDS	276	240	3048	272	224	3184	460	380	4844	348	344	2884
I, ma		110			120			320			180	
Q, ml/min		412			400			692			728	

*In mg/liter, except as otherwise indicated.

TABLE B-4.

RESULTS OF ANALYSES - SYSTEM II, SERIES 1 AND 2

	SERIES 1		SERIES 2							
	1/4/63		1/29/63		2/5/63		2/8/63		2/12/63	
Constituent*	Influent	Effluent	Influent	Effluent	Influent	Effluent	Influent	Effluent	Influent	Effluent
Sodium, Na^+	42.0	19.0	40.0	12.0	22.2	13.2			29.5	20.2
Potassium, K^+	8.0	4.5	5.4	1.5	10.5	5.3				
Ammonium, NH_4^+			14.2	5.6	18.2	8.5	35.8	23.0	18.5	11.8
Calcium, Ca^{+2}	12.4	8.5	11.6	4.4	11.6	3.4			10.1	6.8
Magnesium, Mg^{+2}	0.8	0.8	1.6		0.9	2.0			1.5	0.5
Chloride, Cl^-	44.1	29.6	23.4	10.1	31.6	13.4	54.7	33.3	27.7	16.7
Bicarbonate, HCO_3^-	129.3	54.3	90.3	31.7	110.4	57.3	203.1	136.6	122.0	91.5
Nitrate, NO_3^-	0.8	0.8	1.0	0.9	1.0	0.9			1.5	1.0
Sulfate, $SO_4^=$	26.0	24.0	26.2	9.6	15.0	7.0			14.3	7.5
Phosphate, $PO_4^{\equiv}$	15.5	10.5	5.9	3.6	8.7	6.3			5.0	2.0
Total Ions			219.6		230.1	117.3			230.1	158.0
ABS			0.11	0.09	0.37	0.4			0.09	0.06
COD	240.0	54.7	30.8	27.4	44.7	43.7			5.3	14.0
Alkalinity, meq/l	2.12	0.89	1.48	0.52	1.81	0.94	3.3	2.24	2.0	1.5
pH	6.6	6.4	7.2	6.3	7.6	6.9	7.5	7.3	7.4	7.4
Resistivity, ohm-cm	1850	4300	4190	9180	3550	6700			3550	5000
Temperature, °C			17.0	18.0	19.0	19.5	15.0	15.0	18.0	18.0
Turbidity, SiO_2										
Organic N										
I, ma	115		100		90				59	
Q, ml/min	120-130		120		120				116	

TABLE B-4 (continued)

Constituent*	2/15/63		2/19/63		2/27/63		3/8/63		3/12/63	
	Influent	Effluent	Influent	Effluent	Influent	Effluent	Influent	Effluent	Influent	Effluent
Sodium, Na^+			30.0	26.8			38.0	27.4		
Potassium, K^+			7.0	3.0			7.0	2.0		
Ammonium, NH_4^+	30.4	20.4	23.6	14.9	25.15	20.85	20.0	11.9	15.9	10.8
Calcium, Ca^{+2}			12.1	8.9	14.3	13.2	11.1	11.1	12.45	9.66
Magnesium, Mg^{+2}			0.51	0.37	0.86	0.58	1.5	1.6	2.23	1.50
Chloride, Cl^-	50.0	30.3	28.2	19.4	22.6	19.4	37.0	25.0	29.8	22.2
Bicarbonate, HCO_3^-	184.8	145.8	134.2	113.5	144.0	131.8	137.0	98.8	107.4	93.9
Nitrate, NO_3^-			8.0	8.0	32.5	20.5	0.2	negl.	14.0	15.5
Sulfate, $SO_4^=$			10.0	8.0	10.5	9.5	14.2	13.8	15.0	13.6
Phosphate, $PO_4^{\equiv}$			7.5	5.5	15.0	8.5	22.0	12.5	5.0	3.7
Total Ions			261.1	208.37			288.0	204.1		
ABS			0.1	negl.	0.22	negl.	negl.	negl.	0.04	negl.
COD			20.2	16.7	55.0 59.0	59.0 59.0	86.5 49.2	50.1	12.1 18.9	18.9
Alkalinity, meq/l	3.03	2.39	2.2	1.86	2.36	2.16	2.24	1.62	1.76	1.54
pH	7.5	7.5	7.5	7.4	7.4	7.35	7.4	7.3	7.2	7.5
Resistivity, ohm-cm			3000	3900	3250	3400	3300	4200	3500	5200
Temperature, °C	12.5	14.0	19.0	19.0	17.0	16.0	14.0	13.0	18.0	15.5
Turbidity, SiO_2							42.5	25.0		
Organic N										
I, ma			50		23		34		22	
Q, ml/min			120		125		120		135	

TABLE B-4 (continued)

Constituent*	3/15/63		3/19/63		3/26/63		3/29/63		4/2/63	
	Influent	Effluent	Influent	Effluent	Influent	Effluent	Influent	Effluent	Influent	Effluent
Sodium, Na^+										
Potassium, K^+										
Ammonium, NH_4^+	8.9	8.9	11.0	7.25	14.8	8.7	5.7	3.65	6.4	4.3
Calcium, Ca^{+2}	15.9	8.7								
Magnesium, Mg^{+2}	1.72	2.73								
Hardness, Ca^{+2}			17.3	12.2	16.6	11.8	17.9	13.4	16.6	10.7
Chloride, Cl^-	35.1	28.0	28.8	20.4	26.1	17.9	34.3	23.6	28.2	18.7
Bicarbonate, HCO_3^-	100.0	85.4	101.3	78.7	68.3	55.5	67.5	26.7	54.9	39.7
Nitrate, NO_3^-	11.5	8.5	9.0	11.0	11.5	8.5	10.0	5.5	37.0	35.0
Sulfate, $SO_4^=$	22.0	19.0	20.5	15.0	20.0	17.0	12.0	14.0	14.4	14.0
Phosphate, $PO_4^{\equiv}$	6.0	5.7	2.2	1.5	5.0	3.5	4.5	3.0	2.5	1.0
ABS	0.12	0.02	0.04	0.001	0.05	0.04	0.04	negl.	0.10	0.04
COD	41.8 37.0	17.5	13.3 3.8	26.6 5.7			13.6 15.6	8.2 8.2	31.7 31.7	23.8 23.7
Alkalinity, meq/l	1.64	1.40	1.66	1.29	1.12	0.91	1.106	0.438	0.90	0.65
pH	7.4	7.5	7.5	7.4	6.85	6.85	6.8	6.7	7.1	6.7
Resistivity, ohm-cm	3300	4200	3100	4000	3500	5200	4400	6200	4300	6200
Temperature, °C	18.5	15.0	17.0	17.5	22.5	17.5	16.8	15.5	18.0	18.5
Turbidity, SiO_2			9.0	5.0			25.0	18.0		
Organic N										
I, ma	25		30		31		37		36	
Q, ml/min	125		125		125		125		125	

TABLE B-4 (continued)

Constituent*	4/5/63		4/9/63		4/12/63		4/19/63		4/23/63	
	Influent	Effluent	Influent	Effluent	Influent	Effluent	Influent	Effluent	Influent	Effluent
Sodium, Na^+										
Potassium, K^+										
Ammonium, NH_4^+	9.2	5.8			10.4	4.2	15.8	10.9	13.2	10.05
Hardness, Ca^{+2}	18.0	11.2	15.3	9.6	13.6	9.1	11.48	7.88	14.14	8.8
Chloride, Cl^-	32.0	23.0	27.7	19.9	32.2	22.4	29.6	21.0	28.8	24.4
Bicarbonate, HCO_3^-	92.7	74.4	62.2	43.9	79.9	58.6	91.5	68.9	95.8	78.7
Nitrate, NO_3^-	15.5	12.5	29.5	12.0	9.0	8.5	1.0	0.5	4.5	2.75
Sulfate, $SO_4^=$	15.0	10.0	22.0	20.0	17.0	17.0	32.6	21.2	22.0	16.4
Phosphate, $PO_4^{\equiv}$	12.5	10.0	24.0	21.0	31.5	27.0	29.5	28.5	24.5	20.5
ABS	0.2	0.08	0.06	negl.	0.06	0.02	1.0	0.3	0.042	0.04
COD	62.8 61.2	16.0 16.0	18.15 18.15	29.1 25.4	22.8 20.5	19.7 19.7	34.0 32.2	28.6 28.6	17.5 17.5	15.5 17.5
Alkalinity, meq/l	1.52	1.22	1.02	0.72	1.31	0.96	1.50	[illegible]	1.57	1.29
pH	7.0	6.95	7.2	7.4	6.8	6.8	6.85	6.8	7.4	7.4
Resistivity, ohm-cm	3800	4800	4700	6200	3380	4650	3100	3500	3600	4760
Temperature, °C	18.5	18.0	18.0	16.0	17.2	18.7	20.2	19.6	15.4	14.0
Turbidity, SiO_2					22.0	15.0	14.0	10.0	14.0	10.0
Organic N	1.9	0.6	2.05	0.82	0.78	0.66				
I, mm	40		38		42		36		36	
Q, ml/min	125		123		125		118		112.5	

TABLE B-4 (continued)

Constituent[*]	4/30/63		5/3/63		5/10/63		5/17/63		5/24/63	
	Influent	Effluent	Influent	Effluent	Influent	Effluent	Influent	Effluent	Influent	Effluent
Sodium, Na^+					42.5	37.2	38.4	25.0	40.0	34.0
Potassium, K^+					20.5	17.5	16.9	11.8	16.5	12.5
Ammonium, NH_4^+	9.78	5.85	16.28	16.30	18.18	20.52	13.8	8.7	16.5	12.7
Hardness, Ca^{+2}	10.74	6.08	13.06	6.44	8.22	5.0	11.64	6.26	12.56	7.96
Chloride, Cl^-	22.4	23.6	29.2	43.5	37.8	41.2	25.1	18.1	30.0	20.7
Bicarbonate, HCO_3^-	99.8	54.3	146.4	128.1	142.7	115.7	111.6	92.1	118.3	102.5
Nitrate, NO_3^-	22.0	8.5	5.5	22.0	1.0	0.25	33.5	2.5	5.5	4.5
Sulfate, $SO_4^=$	16.0	11.6			17.0	27.0	23.0	15.0	26.0	16.0
Phosphate, $PO_4^{\equiv}$	23.0	12.0	31.0	23.0	22.75	16.25	40.5	27.5	41.0	36.0
Total Ions					310.7	280.4	314.4	207.0	306.4	246.9
ABS	0.08	0.04	0.02	0.01	negl.	negl.	< 0.01	0.01	< 0.01	< 0.01
COD	5.7 9.5	11.4 7.6	35.2 35.2	27.2 27.2	22.7 25.0	25.0 25.0	39.6 36.6	39.6 38.8	19.5 19.5	21.0 18.0
Alkalinity, meq/l	1.636	0.89	2.40	2.10	2.34	1.896	1.83	1.51	1.94	1.68
pH	7.2	6.5	7.18	6.95	6.5	6.5	7.0	6.8	7.0	7.05
Resistivity, ohm-cm	4000	6600	3100	3200	2800	3200	3500	5000	3000	4000
Temperature, °C	18.5	18.7	19.8	18.1	21.0	20.0	18.8	18.6	18.0	17.0
Turbidity, SiO_2	6.0	4.0			51.0	42.0	14.0	12.0	26.0	20.5
Organic N									2.0	1.6
I, ma	68		59		39		54		50	
Q, ml/min	112		102		133		135		125	

TABLE B-4 (continued)

Constituent*	5/29/63 Influent	5/29/63 Effluent	5/31/63 Influent	5/31/63 Effluent	6/12/63 Influent	6/12/63 Effluent	6/25/63 Influent	6/25/63 Effluent
Sodium, Na^+	47.0	44.0					44.0	
Potassium, K^+	20.0	18.0					20.0	
Ammonium, NH_4^+	19.5	18.9	15.57	14.64	17.9	18.4	20.11	
Hardness, Ca^{+2}	11.9	7.12	10.9	7.7	12.44	8.66	19.0	
Chloride, Cl^-	28.0	23.8	27.3	22.7	26.8	25.8	28.8	
Bicarbonate, HCO_3^-	156.2	141.5	139.1	125.0	126.3	115.3	26.8	
Nitrate, NO_3^-	9.0	6.0	2.0	0.5			15.0	
Sulfate, $SO_4^=$	21.0	19.0	22.0	18.2			160.0	
Phosphate, $PO_4^{\equiv}$	30.0	27.5	30.5	26.0			39.0	
Total Ions	342.6	305.8					372.7	
ABS	<0.1	<0.1	<0.1	<0.1				
COD	38.8 40.7	40.0 40.0	41.0 44.0	37.4 32.3			20.2 23.5	
Alkalinity, meq/l	2.56	2.32	2.28	2.05	2.07	1.89	0.44	
pH	7.2	7.2	7.4	7.45	7.4	7.25	6.2	
Resistivity, ohm-cm	2500	3000	2500	3100	3500	3300		
Temperature, °C	19.5	19.0	24.0	21.8	16.0	16.8		
Turbidity, SiO_2	11.0	11.0	26.0	24.8			39.0	
Organic N	2.96	2.88	3.1	3.1				
I, ma	46		37		29		360	
Q, ml/min	122		127		125		380	

*In mg/liter, unless otherwise indicated.

TABLE B-5.

RESULTS OF ANALYSES - YSTEM II, PRODUCT/WASTE RATIO STUDY

	6/26/63			6/28/63			7/2/63			7/8/63		
Constituent*	Influent	Effluent	Conc	Influent	Effluent	Conc	Influent	Effluent	Conc	Influent	Effluent	Conc
Sodium, Na^+		25.0	400.0	70.0	65.0	250.0	36.0	31.0	88.0	34.0	26.0	500.0
Potassium, K^+		8.0	202.0	12.8	11.2	32.0	51.0	44.0	174.0	20.0	15.0	470.0
Ammonium, NH_4^+				23.82	9.45	30.8	14.7	8.8	46.1	16.9	11.8	384.0
Hardness, Ca^{+2}		10.8	182.8	11.3	8.1	55.6	14.6	11.6	59.6	11.88	8.54	521.2
Chloride, Cl^-		21.4	40.1	21.2	17.4	78.2	53.7	41.2	645.0	61.0	54.5	3980.0
Bicarbonate, HCO_3^-				76.9	68.9	45.1	109.2	91.5	0	80.5	43.9	0
Nitrate, NO_3^-		22.0	385.0	34.0	32.0	180.0	8.4	6.7	25.0	<2.5	<2.5	20.0
Sulfate, $SO_4^=$		100.0	1300.0	85.0	63.0	750.0	33.8	30.2	82.0	14.8	13.2	250.0
Phosphate, $PO_4^{\equiv}$		30.0	220.0	25.0	20.0	120.0	23.0	22.0	60.0	9.2	9.0	126.0
Total Ions				360.0	295.1	1541.7	344.4	287.0	1179.7	250.8	184.4	6251.2
ABS			36.75				3.6	0.6	0.6	<0.1	<0.1	0.4
COD		26.8 26.8	43.5 42.0									
Alkalinity, meq/l				1.26	1.13	0.74						
pH		3.2	2.7	6.7	6.7		8.3	8.0	2.3	7.2	6.5	
Resistivity, ohm-cm				2700	3200		1970	2600		2500	3300	
Temperature, °C				29.3	30.0							
Turbidity, SiO_2				15.0	15.0		51.0	70.0	200.0			
Organic N												
I, ma					78			95			100	
Q, ml/min					350			330			460	

TABLE B-5 (continued)

Constituent*	7/12/63			7/16/63			7/19/63			7/23/63		
	Influent	Effluent	Conc	Influent	Effluent	Conc	Influent	Effluent	Conc	Influent	Effluent	Conc
Sodium, Na^+	86.0	75.0	2450.0	66.5	53.5	1650.0	50.0	43.0	1800.0	66.0	54.0	2050.0
Potassium, K^+	46.0	39.5	1600.0	25.0	21.6	875.0	18.0	14.0	765.0	26.5	18.0	750.0
Ammonium, NH_4^+	22.75	17.6	675.0	21.1	19.0	640.0	26.43	14.82	778.0	23.65	19.75	770.0
Hardness, Ca^{+2}	11.26	8.08	600.0	11.42	9.02	1936.1	14.8	10.8	816.0	9.0	6.4	1180.0
Chloride, Cl^-	110.0	105.0	6800.0	58.8	55.5	5300.0	65.7	58.6	6800.0	21.2	19.7	8000.0
Bicarbonate, HCO_3^-	107.0	83.0	0	145.0	113.0	0	116.5	75.0	0	134.2	106.1	0
Nitrate, NO_3^-	42.0	32.5	1280.0	10.0	6.5	220.0	<2.5	<2.5	475.0	5.0	<2.5	80.0
Sulfate, $SO_4^=$	26.7	25.2	445.0	23.7	20.7	550.0	29.0	25.0	900.0	44.0	37.0	1175.0
Phosphate, $PO_4^{\equiv}$	34.0	29.0	420.0	18.5	17.0	274.0	20.0	19.0	395.0	19.0	11.0	265.0
Total Ions	485.7	414.9	14270	380.0	315.8	11445.1	342.9	262.7	12629	348.6	274.5	14270
ABS	1.2	0.8	0.6	negl.	negl.	0.4	negl.	negl.	0.3	negl.	negl.	0.4
COD	36.7 36.4	34.4 34.5	46.3 200.0	18.2 17.9	14.1 15.2		14.1 14.1	12.9 14.1		6.85 8.75	6.85 6.85	
Alkalinity, meq/l							1.91	1.23	0	2.20	1.74	
pH	7.15	7.0	2.2	7.6	7.3	2.4	7.3	6.7	1.7	7.2	6.9	2.1
Resistivity, ohm-cm	1280	1490	39	1800	2100	52	1900	2400	36	1800	2800	39
Temperature, °C	16.0	18.3	18.1	28.0	27.4	28.5	29.0	31.5	31.5	22.4	24.0	24.5
Turbidity, SiO_2	12.5	12.5		27.5	11.0	19.0	19.0	20.0	1.5	33.0	33.0	35.0
Organic N	3.25	3.55		1.42	1.73	56.5	1.1	1.3		1.5	1.4	48.5
TDS	400	380	15708	168	168	10056	224	160	10150	448	380	14944
I, ma		260			150			240			420	
Q, ml/min		485			400			500			720	

TABLE B-5 (continued)

	7/26/63			7/29/63		
Constituent*	Influent	Effluent	Conc	Influent	Effluent	Conc
Sodium, Na^+	85.0	75.0	3200	65.0	60.0	4000
Potassium, K^+	15.0	12.0	1700	8.5	6.0	1700
Ammonium, NH_4^+	15.22	12.8	1150	11.0	6.75	1390
Hardness, Ca^{+2}	12.0	8.2	1160	10.7	6.0	1640
Chloride, Cl^-	72.3	59.4	9450	58.5	48.7	12180
Bicarbonate, HCO_3^-	76.1	52.2	0	42.09	32.94	0
Nitrate, NO_3^-	20.0	18.0	750	30.0	20.0	240
Sulfate, $SO_4^=$	20.0	19.0	1300	19.0	18.0	2300
Phosphate, $PO_4^{\equiv}$	14.0	13.0	300	18.0	16.0	250
Total Ions	329.6	269.6	19010	252.8	214.4	23700
ABS				negl.	negl.	0.25
COD	6.6[illegible] 8.75	6.85 6.85		14.4 14.8	14.4 13.7	
Alkalinity, meq/l				0.69	0.54	0
pH	7.0	6.8	5.0	6.5	5.8	3.8
Resistivity, ohm-cm	1700	2350	29	2300	3500	22
Temperature, °C	33.4	36.0	35	34.5	37.4	38
Turbidity, SiO_2	6.5	6.5	24.5	24.9	22.5	40
TDS				228.0	184.0	24124
Organic N	3.1	1.4	26.5	1.71	1.69	
I, ma		270			310	
Q, ml/min		680			550	

*In mg/liter, unless otherwise indicated.

APPENDIX C

Voltage Probe Data

TABLE C-1(1).

VOLTAGE PROBES - SYSTEM I

				CELL NUMBER													
Date	Hrs.	E_T V	Cath. volts	1 Ω	C Ω	2 Ω	C Ω	3 Ω	C Ω	4 Ω	C Ω	5 Ω	An volts	R_{in} Ω-cm	R_{out} Ω-cm	n ml/min	I ma.
12/30		5.0	0.5	6.5	0	9.7	1.6	6.5	0	4.8	3.2	16.1	1.7		9270	120	62
12/30		25.0	1.0	22.3	0	27.0	0	30.0	0	27.0	0	46.6	1.0	6170	19120	120	150
1/2		25.0	1.0	30.2	0	34.5	0	34.5	0	43.1	0	51.6	2.0		14400		116
2/11	117	16.5	1.4	16.8	0	17.9	0	20.0	0	18.9	3.2	61.0	2.0	3550	9900		95
2/18	213	15.5	2.0	16.1	0	15.1	6.4	9.7	0	14.0	5.4	54.8	2.2	3100	5780	120	93
2/20	254	10.2	1.9	7.1	2.4	26.1	0	14.3	4.8	16.7	11.9	102.3	0.5	3000	4600	115	42
2/26	305	5.7	1.1	10.3	0	10.3	0	10.3	0	17.2	76.0	0	1.0	2970	3900	115	29
2/27	338	5.9	1.2	11.5	0	7.7	0	15.4	0	15.4	11.5	50.0	1.8	2900	3700	112.5	26
3/7	386	6.6	1.4	5.3	0	5.3	0	13.1	0	7.9	5.3	2.6	3.7	3600	4700	115	38
3/8	406	7.0	1.1	16.7	0	6.7	0	13.3	0	13.3	0	13.3	4.0	3300	4400	115	30
3/11	432	4.7	1.6	3.3	0	3.3	0	3.3	0	3.3	6.7	6.7	2.4	2700	3550	120	30
3/12	452	5.6	1.1	10.0	0	15.0	0	15.0	0	15.0	10.0	5.0	3.1	3000	4800	113	20
3/13	480	5.9	1.2	13.6	0	9.1	0	13.6	0	18.2	9.1	13.6	3.0	3500	4500		22
3/20	576	6.2	1.1	10.0	0	10.0	0	13.3	0	10.0	3.3	6.7	3.5	3200	4300	102	30
3/25	632	4.1	0.7	18.5	0	3.7	0	7.4	0	7.4	3.7	14.8	1.9	3100	6200	107	27
3/26	656	4.2	0.8	12.5	0	6.3	0	6.3	0	6.3	3.1	12.5	1.9	3600	7000		32
3/29	706	4.8	1.1	11.8	0	2.9	0	5.9	0	5.9	0	26.5	1.9	4000	7800	105	34
4/1	728	3.8	1.0	4.2	0	2.1	0	2.1	2.1	2.1	4.2	2.1	1.9	1700	2400	105	48
4/2	757	6.4	0.6	27.0	2.7	5.4	2.7	16.2	0	8.1	2.7	40.6	1.9	4000	8000	105	37
4/3	774	6.6	1.0	5.6	22.2	5.6	0	16.7	0	8.3	2.8	41.7	1.9	4300	8000		36

TABLE C-1(1) (continued)

Date	Hrs.	E_T V	Cath volts	CELL NUMBER									An volts	R_{in} Ω -cm	R_{out} Ω -cm	Q ml/min	I ma.
				1 Ω	C Ω	2 Ω	C Ω	3 Ω	C Ω	4 Ω	C Ω	5 Ω					
4/4	780	4.6	0.5	11.6	0	4.6	0	7.0	0	7.0	0	20.9	1.9	3200	3900	105	43
4/6	820	7.2	1.2	22.8	0	5.7	5.7	17.2	0	5.7	2.9	54.3	2.0			105	35
4/8	828	7.2	0.6	31.6	0	10.5	0	21.1	0	7.9	2.6	50.0	1.9	4000	8600	90	38
4/9	857	7.8	1.4	29.4	0	5.9	0	23.5	0	11.8	5.9	55.9	1.9	4200	6800	115	34
4/11				Bad probe but system shows no conc. resistances													
4/12				Bad probe but system shows no conc. resistances													
4/19	947	8.9	0.7	28.0	0	8.0	4.0	26.0	0	10.0	4.0	46.0	2.0	3100	5100	130	50
4/23	980			Bad probe but system shows no conc. resistances										3600	5000	125	36
4/25	998	4.6	1.0	0	4.2	2.1	0	4.2	2.1	10.6	4.2	0	2.3	3900	5800		47
4/26	1020	5.7	1.1	6.7	0	6.7	2.2	11.1	0	4.4	0	2.2	3.1	3400	5800	125	45
4/29	1044	5.7	1.0	8.7	0	10.9	0	13.0	0	6.5	0	21.7	1.9	4000	6600	125	46
5/2	1086	4.9	0.7	7.1	0	7.1	4.8	9.5	0	7.1	0	16.7	2.0	2900	4700	125	42
5/9	1139	3.0	0.4	13.2	0	2.6	0	5.2	0	3.9	0	2.6	1.55	2300	3200	127	38
5/10	1170	3.9	0.9	4.8	0	4.8	0	4.8	0	4.6	0	33.3	0.8	2400	3600		42
5/11	1188	4.2	0.7	11.8	0	8.8	0	5.9	0	8.8	0	17.6	1.7	3950	4600	130	34
5/14	1218	4.3	0.7	10.5	0	5.3	0	7.9	0	5.3	0	23.7	1.6	2700	4000	125	38
5/17	1242	5.2	1.3	4.6	0	7.0	0	9.3	0	7.0	0	25.6	1.6	3000	4000	130	43
5/21	1271	4.8	1.0	5.0	0	7.5	0	7.5	0	5.0	0	25.0	1.8	3290	4000	125	40
5/22	1290	5.3	1.2	8.6	0	5.7	0	11.4	0	5.7	0	34.3	1.8	4400	5200	125	35
5/25	1342	6.9	1.1	21.0	0	13.2	0	15.8	0	10.5	0	39.5	2.0	3100	4100	125	38
5/28	1370	6.3	1.3	10.4	0	6.2	0	10.4	0	4.2	0	29.2	2.1	2400	3000	125	48

TABLE C-1(2).

VOLTAGE PROBES - SYSTEM I

					CELL NUMBER													
Date	E_T V	I ma.	Q ml/min	Cath V	C Ω	1 Ω	C Ω	2 Ω	C Ω	3 Ω	C Ω	4 Ω	C Ω	5 Ω	C Ω	Anode V	R_{in} Ω-cm	R_{out} Ω-cm
7/1	5.2	100	402	0.2	6.0	4.0	0	4.0	0	6.0	0	3.0	1.0	2.0	1.0	2.3		
7/3	5.5	110	340	0.3	1.8	7.3	0	4.5	0	2.7	0	3.6	0	4.5	2.7	2.2	1950	2600
7/6	5.8	105	415	1.4	2.9	2.9	0	2.9	0	4.8	0	1.9	0	3.8	1.9	2.2	2300	3000
7/8	4.8	90	460	0.8	1.1	3.3	0	3.3	0	3.3	0	4.4	0	2.2	1.1	2.3	2400	2700
7/9	4.9	200	385	0.8	1.0	1.5	0	1.0	0	1.0	0	1.0	0	1.5	0	2.7	1200	1400
7/10	4.8	250	480	0.5	1.2	1.2	0	0.8	0	0.8	0	0.8	0	1.6	0	2.7	600	900
7/11	4.9	120	500	0.7	0	3.3	0	2.5	0	2.5	0	2.5	0	3.3	0	2.5	1200	1500
7/12	4.4	160	480	0.5	0.6	1.9	0	1.2	0	1.2	0	1.2	0	2.5	0	2.2		
7/15	4.9	130	470	0.6	0.8	2.3	0	3.1	0	2.3	0	2.3	0	3.1	0.8	2.4		1400
7/16	4.6	120	400	0.5	0.8	2.5	0	2.5	0	2.5	0	1.7	0	4.2	0.8	2.3	1600	1850
7/17	5.0	160	420	0.5	0.6	1.9	0	2.5	0	1.9	0	2.5	0	3.1	0.6	2.4	1350	1720
7/18	5.4	100	400	0.7	0	5.0	0	4.0	0	5.0	0	5.0	0	3.0	1.0	2.4	1900	2300
7/19	6.7	120	350	0.8	0.8	5.8	0	4.2	0	4.2	0	4.2	0	6.7	0.8	2.7		
7/22	6.6	140	550	0.6	0.7	5.0	0	5.0	0	3.6	0	4.3	0	5.0	0.7	2.6	1920	2300
7/23	7.8	320	680	0.6	0.6	1.9	0	1.9	0	4.1	0	2.8	0	2.5	0.3	2.7		
7/24	8.2	180	840	0.4	1.1	2.2	0	6.1	0	6.1	0	5.5	0	7.2	0.5	2.6		
7/25	8.0	160	720	0.6	3.7	1.3	0	6.2	0	7.5	0	2.5	0	6.2	2.5	2.6	1900	2100
7/26	18.0	300	720	2.0	3.3	3.3	0	10.0	0	10.0	0	3.3	0	13.3	3.3	3.0	1600	2400
7/26	4.7	210	720	0.2	0.5	1.9	0	1.9	0	1.4	0	1.4	0	1.9	0.9	2.4		
7/27	7.2	150	920	0.6	0.7	3.3	0	6.7	0	5.3	0	2.7	0	8.0	1.3	2.4	1850	2100
7/29	7.2	180	728	0.4	0.5	3.9	0	4.4	0	4.4	0	3.3	0	6.7	0.5	2.5	1150	1950

TABLE C-2(1).

VOLTAGE PROBES - SYSTEM II

Date	Hrs.	E_T V	Cath volts	CELL NUMBER									An volts	R_{in} Ω-cm	R_{out} Ω-cm	Q ml/min	I ma.
				1 Ω	C Ω	2 Ω	C Ω	3 Ω	C Ω	4 Ω	C Ω	5 Ω					
1/2		25.0	1.0	23.0	0	30.8	0	30.8	0	30.8	0	53.8	2.0		≥19000		130
2/11	120	15.3	1.1	15.4	10.7	10.7	18.5	16.9	30.8	15.4	27.7	26.2	3.0	3550	5600		65
2/18	215	15.0	1.4	35.0	0	25.0	0	25.0	25.0	28.3	3.3	5.0	4.8	3500	5000	120	60
2/20	256	13.2	1.2	9.6	0	5.8	23.1	5.8	67.3	3.8	61.5	5.8	2.5	3000	3750	130	52
2/26	308	12.0	1.0	22.7	0	0	81.9	31.8	104.5	13.6	9.1	109.0	2.8	3050	3400	125	22
2/27	340	12.0	1.7	4.5	0	4.5	50.0	13.6	136.3	18.2	136.3	13.6	2.0	3250	3400	125	22
3/7	388	3.6	0.5	9.1	0	6.1	0	6.1	0	6.1	0	9.1	1.9	3700	4600	120	33
3/8	408	3.6	0.4	8.8	0	5.9	0	5.9	0	8.8	0	8.8	1.9	3300	4200	120	34
3/11	434	3.6	0.4	10.5	0	5.3	0	5.3	0	5.3	0	10.5	1.8	2700	3520	123	38
3/12	454	3.5	0.4	10.7	0	7.2	0	7.2	0	7.2	0	14.3	1.8	3500	5200	135	28
3/13	482	3.5	0.4	9.7	0	6.5	0	6.5	0	6.5	0	12.9	1.8	3550	4700		31
3/20	578	2.8	0.3	10.0	0	5.0	0	5.0	0	5.0	0	10.0	1.8	3150	3700	125	20
3/25	634	4.2	0.6	11.1	0	11.1	0	11.1	0	11.1	0	22.2	1.8	5200	6800	120	27
3/26	658	4.0	0.5	9.7	6.5	3.2	0	9.7	0	6.5	0	19.4	1.8	3500	5200		31
3/29	708	4.9	0.6	7.7	2.6	7.7	2.6	12.8	0	10.2	0	17.9	1.9	4400	6200	125	39
4/1	730	3.8	0.7	4.1	0	4.1	0	6.1	0	4.1	0	8.2	1.8	1750	2700	125	49
4/2	759	5.1	0.7	11.1	0	11.1	0	13.9	0	13.9	0	19.4	1.9	4300	6200	125	36
4/3	776	5.0	0.7	11.1	0	11.1	0	13.9	0	11.1	0	16.7	2.0	4300	6600		36
4/4	780	4.4	0.6	8.7	0	4.3	0	8.7	0	6.5	0	10.9	2.0	3000	3900	125	46
4/6	822	4.6	0.5	12.5	5.0	0	15.0	7.5	0	5.0	0	10.0	1.9				40

TABLE C-2(1) (continued)

				CELL NUMBER													
Date	Hrs.	E_T V	Cath volts	1 Ω	C Ω	2 Ω	C Ω	3 Ω	C Ω	4 Ω	C Ω	5 Ω	An volts	R_{in} Ω -cm	R_{out} Ω -cm	Q ml/min	I ma.
4/8	830	5.4	0.7	11.1	0	11.1	0	11.1	0	11.1	0	20.0	1.8	3100	5600	122	45
4/9	859	5.8	0.8	13.2	0	15.8	0	10.5	0	13.2	0	26.3	1.9	4700	6200	123	38
4/12	901	5.2	0.8	10.0	0	10.0	0	12.5	0	10.0	0	17.5	2.0	3800	5000	125	40
4/19	949	4.2	0.9	6.4	0	4.2	0	4.2	0	4.2	0	10.6	1.9	3100	4600	125	47
4/23	982	5.3	1.1	8.5	0	10.6	0	6.4	0	8.5	0	14.9	1.9	3900	4700	125	47
4/25	999	4.2	0.7	4.8	0	4.8	0	4.8	0	4.8	0	4.8	2.0	3900	6000		62
4/26	1023	4.7	1.1	5.0	1.7	5.0	0	5.0	0	3.3	0	6.7	2.0	3600	4600	113	60
4/29	1045	4.7	1.2	2.9	0	4.3	0	4.3	0	4.3	0	5.8	2.0	3500	5400	110	69
5/2	1090	4.2	1.0	3.8	0	3.8	0	5.8	0	3.8	0	5.8	2.0	3100	4400	115	52
5/9	1141	3.6	1.1	5.0	0	4.0	0	4.0	0	4.0	0	4.0	1.5	2300	3500	130	50
5/10	1171	3.5	1.0	7.0	0	2.3	0	4.6	0	4.6	0	4.6	1.5	3100	3900		43
5/11	1190	3.9	1.2	7.5	0	5.0	0	7.5	0	5.0	0	5.0	1.5	3950	4600	130	40
5/14	1219	3.6	0.9	10.7	7.1	3.6	0	10.7	0	3.6	0	3.6	1.6	2700	3000		28
5/17	1243	4.3	1.3	6.7	2.2	4.4	0	4.4	0	4.4	0	6.7	1.7	3300	3900		45
5/21	1272	5.4	1.5	7.5	12.5	0	7.5	5.0	5.0	5.0	5.0	5.0	1.8	3000	3600	130	40
5/22	1290	4.6	1.2	9.5	9.5	2.4	0	4.8	0	4.8	0	4.8	1.9	3800	4500	132	42
5/25	1343	6.2	1.3	9.4	15.1	9.4	0	5.7	0	5.7	0	11.3	2.1	3100	4600	125	53
5/28	1371	5.8	1.2	3.9	25.5	5.9	0	5.9	0	3.9	0	5.9	2.1	2350	3000	125	51

TABLE C-2(2).

VOLTAGE PROBES - SYSTEM II

Date	E_T V	I ma.	Q ml/min	Cath V	CELL NUMBER											Anode V	R_{in} Ω-cm	R_{out} Ω-cm
					C Ω	1 Ω	C Ω	2 Ω	C Ω	3 Ω	C Ω	4 Ω	C Ω	5 Ω	C Ω			
6/25	11.0	420	435	2.0	1.9	2.4	0	1.9	0	2.6	0.9	2.1	0	0.9	0.9	2.2		
6/28	5.75	25	410	0.9	4.0	16.0	4.0	12.0	20.0	12.0	32.0	0	16.0	12.0	8.0	1.5		
7/1	5.8	117	355	0.4	1.7	5.1	0	3.4	0	5.1	0	4.3	0	4.3	1.7	2.4		
7/3	6.0	110	275	0.2	3.6	7.3	0	2.7	0	4.5	0	5.4	0	6.4	0.9	2.4	1800	2400
7/6	5.8	110	460	0.3	0.9	5.4	0	4.5	0	3.6	0	5.4	0	4.5	1.8	2.6	2500	3300
7/8	5.0	95	480	0.8	2.1	4.2	0	2.1	0	3.2	0	3.2	0	3.2	1.0	2.4	2500	3350
7/9	5.0	230	415	0.6	0.9	2.2	0	1.3	0	0.9	0	1.3	0	0.9	0	2.7	1000	1200
7/10	4.9	120		0.9	0.8	3.3	0	1.7	0	1.7	0	2.5	0	1.7	0	2.6		
7/11	5.2	110		1.2	0.9	3.6	0	1.8	0	1.8	0	2.7	0	2.7	0.9	2.5		
					Added more acid to cathode stream													
7/11	5.2	200		0.5	0.5	3.0	0	1.5	0	1.5	0	1.0	0	2.0	0.5	2.7	1600	1820
7/12	4.8	240	480	0.4	0.4	1.7	0	2.1	0	0.4	0	1.2	0	1.7	0	2.6		
7/15	5.4	170	550	0.7	0	2.9	0	2.3	0	1.8	0	2.9	0	1.8	0.6	2.6	1930	2250
7/16	5.4	160	450	0.7	0.6	3.1	0	2.5	0	1.9	0	1.2	0	1.3	0.6	2.6	1600	1900
7/17	5.2	200	460	0.8	0	2.0	0	1.5	0	2.0	0	1.5	0	1.5	0.5	2.6	1400	1680
7/18	5.4	140	460	1.1	0.7	2.9	0	2.1	0	1.4	0	2.9	0	2.1	0.7	2.5	1700	1900
7/19	7.7	250	460	0.5	0.8	3.6	0	2.4	0	4.0	0	2.0	0	3.6	0.8	2.9		
7/22	7.8	170	420	0.6	0	7.1	0	2.3	0	7.1	0	2.3	0	4.7	1.2	3.0	2100	2300
7/22	7.8	200	720	0.6	0	2.0	1.0	2.0	0	2.5	0	2.5	1.0	3.0	2.0	4.0	2100	2300
7/23	12.0	405	680	2.0	0	4.9	0	2.0	0	2.0	0	3.9	0	4.9	0.5	3.6	1800	2800

TABLE C-2(2) (continued)

Date	E_T V	I ma.	Q ml/min	Cath V	CELL NUMBER											Anode V	R_{in} Ω-cm	R_{out} Ω-cm
					C Ω	1 Ω	C Ω	2 Ω	C Ω	3 Ω	C Ω	4 Ω	C Ω	5 Ω	C Ω			
7/24	9.5	320	680	0.3	0.6	4.7	0	2.8	0	3.7	0	4.4	0	4.1	0.3	2.6		
7/25	9.8	290	720	0.4	0.7	4.8	0	4.8	0	3.4	0	4.5	0	4.1	0.3	2.8	2200	3000
7/26	12.0	240	360	0.5	2.1	8.3	0	8.3	0	4.2	0	4.2	0.4	7.1	0.8	3.0	1900	2350
					Added HCl to conc. stream to reduce conc. pH from 6.6 to 2.0													
7/26	12.0	430		0.5	0	3.5	0	3.5	0	3.5	0	3.5	0	3.5	1.2	3.5		
7/27	12.0	310	680	0.1	0	6.1	0	6.4	0	5.8	0	5.5	0	4.8	1.6	2.5	2500	3600
7/29	11.5	320	550	0.5	0	4.7	0	4.7	0	3.1	0	6.2	0	4.7	1.6	3.0	2300	3500

APPENDIX D

Polarization Runs

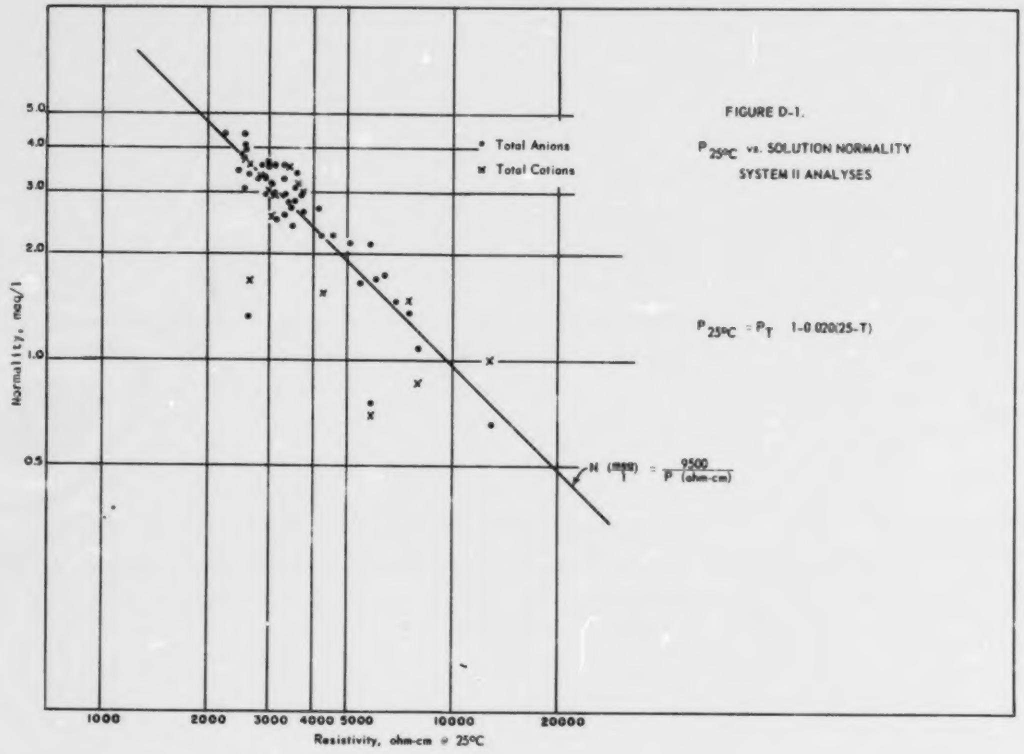

FIGURE D-1.

$P_{25°C}$ vs. SOLUTION NORMALITY
SYSTEM II ANALYSES

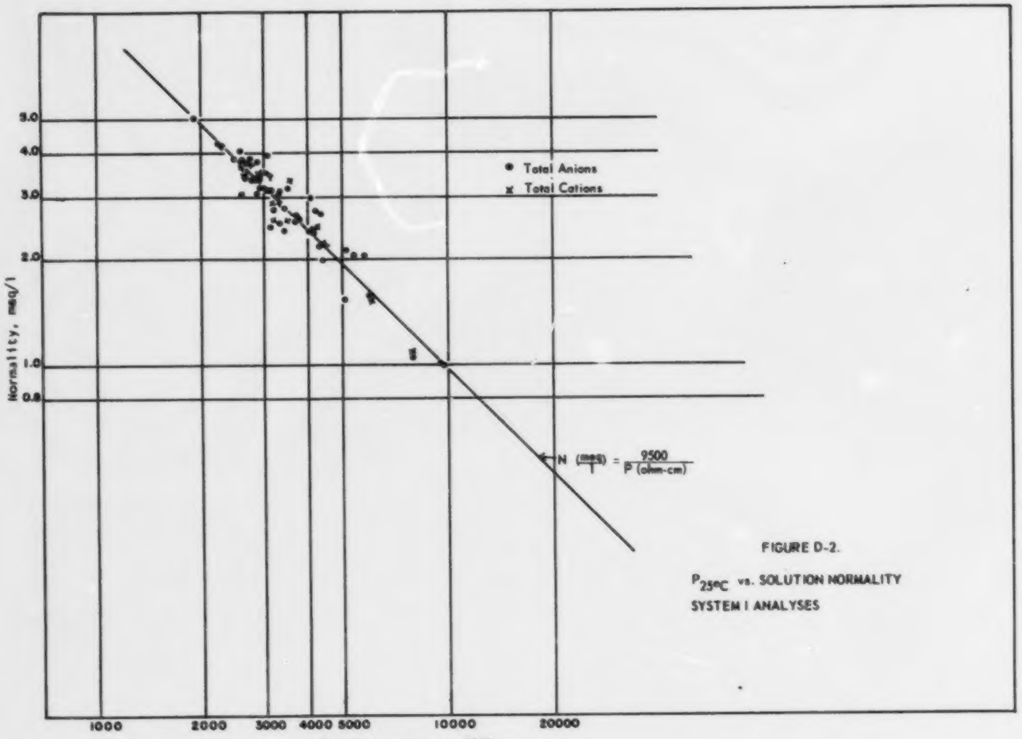

FIGURE D-2.

$\rho_{25°C}$ vs. SOLUTION NORMALITY
SYSTEM I ANALYSES

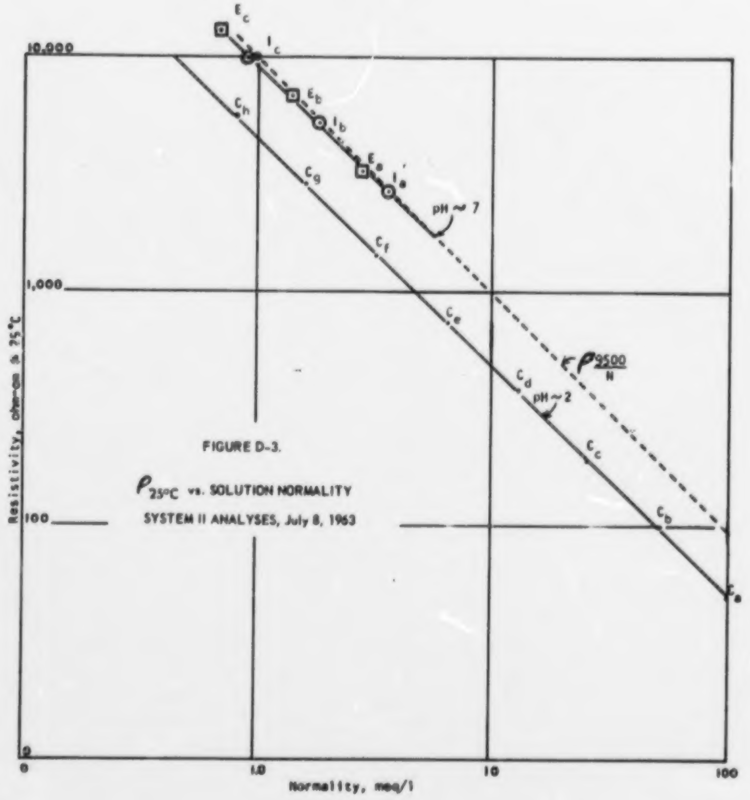

FIGURE D-3.

$\rho_{25°C}$ vs. SOLUTION NORMALITY

SYSTEM II ANALYSES, July 8, 1963

TABLE D-1.

RESISTANCE AS A FUNCTION OF CONCENTRATION

SYSTEM II - July 8, 1963

	P (ohm-cm)	T (°C)	Dilution	Meq/liter
Influent Sample				
(A)	2690	25.2	0	3.53
(B)	5300	25.5	1/2	1.77
(C)	10040	25.6	1/4	0.88
Effluent Sample				
(A)	3300	25.0	0	2.68
(B)	6900	25.3	1/2	1.34
(C)	13040	25.6	1/4	0.67
Conc Sample				
(A)	50	25.5	0	100.0
(B)	98	25.5	1/2	50.0
(C)	190	25.6	1/4	25.0
(D)	380	25.0	1/8	12.6
(E)	739	25.1	1/16	6.3
(F)	1435	25.2	1/32	3.1
(G)	2890	25.4	1/64	1.56
(H)	5600	25.8	1/128	0.78
(I)	10060	25.9	1/256	0.39

APPENDIX E

Selectivity Study Data

TABLE E-1.

MOBILITY RATIO MEASUREMENTS

Membrane Type: AR110DYG-067
Batch No.: 347XA

Water Content: 32.7%
Capacity: 2.58 meq/g
Resistivity: 12.3 ohm-cm^2

Solution	Sample 1			Sample 2		
	R, ohms	d_i, cm	$\frac{R_i}{(100\ d_i)^2}$	R, ohms	d_i, cm	$\frac{R_i}{(100\ d_i)^2}$
(Cl^-)	125	0.0615	3.31	132	0.0612	3.52
	118	0.0608	3.20	132	0.0600	3.66
	120	0.0612	3.21	129	0.0600	3.58
	119	0.0609	3.22	131	0.0605	3.58
(NO_3^-)	165	0.0618	4.31	165	0.0629	4.18
	165	0.0608	4.47	160	0.0613	4.25
	165	0.0608	4.47	168	0.0610	4.51
($SO_4^=$)	270	0.0605	7.38	265	0.0579	7.91
	275	0.0591	7.88	275	0.0602	7.58
	250	0.0580	7.45	255	0.0608	6.89
	275	0.0605	7.51	275	0.0572	8.39
($H_2PO_4^-$)	508	0.0602	13.98	500	0.0585	14.61
	500	0.0580	14.89	498	0.0609	13.41
	500	0.0615	13.25	500	0.0618	13.08
($HPO_4^=$)	330	0.0629	8.33	328	0.0610	8.83
	330	0.0610	8.87	330	0.0628	8.37
	330	0.0610	8.87	335	0.0610	9.02
($PO_4^{\equiv}$)	282	0.0642	6.83	320	0.0640	7.81
	270	0.0615	7.14	300	0.0625	7.68
	280	0.0619	7.31	300	0.0647	7.18

TABLE E-1 (continued)

MOBILITY RATIO MEASUREMENTS

Membrane Type:	AR110CZG-067	Water Content:	40.6%
Batch No.:	236	Capacity:	1.8 meq/g
		Resistivity:	26 ohm-cm^2

Solution	Sample 1			Sample 2		
	R, ohms	d_i, cm	$\frac{R_i'}{(100\, d_i)^2}$	R, ohms	d_i, cm	$\frac{R_i}{(100\, d_i)^2}$
(Cl^-)	79	0.0665	1.79	80	0.0671	1.77
	80	0.0631	1.72	80	0.0689	1.69
	80	0.0700	1.63	80	0.0705	1.61
(NO_3^-)	90	0.0592	2.56	95	0.0689	2.01
	90	0.0608	2.43	95	0.0682	2.04
	110	0.0650	2.60	95	0.0665	2.15
	90	0.0660	2.06			
	92	0.0645	2.21			
($SO_4^=$)	138	0.0562	4.37	137	0.0630	3.45
	135	0.0560	4.31	140	0.0615	3.70
	135	0.0562	4.28	136	0.0610	3.65
($H_2PO_4^-$)	351	0.0596	9.87	360	0.0645	8.67
	350	0.0592	9.99	362	0.0642	8.78
	350	0.0582	10.33	375	0.0650	8.86
($HPO_4^=$)	220	0.0660	5.05	170	0.0589	4.90
	199	0.0662	4.55	170	0.0625	4.35
	198	0.0668	4.43	170	0.0589	4.90
($PO_4^{\equiv}$)	168	0.0635	4.16	150	0.0562	4.75
	162	0.060	4.51	135	0.0555	4.39
	162	0.0582	4.78	135	0.0556	4.38
	160	0.0569	4.95	134	0.0562	4.24

TABLE E-1 (continued)

MOBILITY RATIO MEASUREMENTS

Membrane Type: AR111BZL-183
Batch No.: 184-A

Water Content: 45.2%
Capacity: 1.8 meq/g
Resistivity: 15.2 ohm-cm^2

Solution	Sample 1			Sample 2		
	R, ohms	d_i, cm	$\frac{R_i}{(100\ d_i)^2}$	R, ohms	d_i, cm	$\frac{R_i}{(100\ d_i)^2}$
(Cl^-)	80	0.0578	2.40	72	0.0570	2.22
	72	0.0582	2.12	75	0.0571	2.29
	75	0.0596	2.11	76	0.0572	2.33
	75	0.0585	2.19			
	75	0.0581	2.22			
(NO_3^-)	102.0	0.0639	2.49	96	0.0520	3.55
	101.2	0.0639	2.48	95	0.0518	3.43
	102.0	0.0639	2.49	96	0.0517	3.58
$(SO_4^=)$	128	0.0569	3.96	110	0.0605	3.01
	112	0.0552	3.68	107	0.0602	2.96
	112	0.0542	3.81	115	0.0589	3.31
$(H_2PO_4^-)$	200	0.0625	5.12	204	0.0575	6.16
	190	0.0612	5.08	200	0.0560	6.38
	196	0.0591	5.60	200	0.0540	6.85
$(HPO_4^=)$	135	0.0629	3.41	135	0.0632	3.38
	134	0.0621	3.47	135	0.0625	3.46
	130	0.0635	3.22	125	0.0619	3.27
	132	0.0631	3.32	128	0.0622	3.31
				130	0.0625	3.32
$(PO_4^{\equiv})$	151	0.0649	3.59	120	0.0645	2.88
	120	0.0647	2.87	118	0.0652	2.77
	120	0.0646	2.87	120	0.0660	2.76
	115	0.0647	2.74	119	0.0649	2.83

TABLE E-1 (continued)

MOBILITY RATIO MEASUREMENTS

Membrane Type:	AR111GWL-183	Water Content:	53%
Batch No.:	Special	Capacity:	1.32 meq/g
		Resistivity:	23.3 ohm-cm^2

Solution	Sample 1			Sample 2		
	R, ohms	d_i, cm	$\frac{R_i}{(100\ d_i)^2}$	R, ohms	d_i, cm	$\frac{R_i}{(100\ d_i)^2}$
(Cl^-)	98	0.0589	2.83	95	0.0655	2.12
	95	0.0592	2.71	95	0.0680	2.06
	95	0.0600	2.64	95	0.0723	1.82
(NO_3^-)	125	0.0720	2.42	122	0.0600	3.29
	122	0.0660	2.80	122	0.0612	3.26
	122	0.0680	2.63	125	0.0610	3.28
$(SO_4^{=})$	132	0.0600	3.67	130	0.0539	4.48
	123	0.0592	3.51	129	0.0545	4.34
	122	0.0585	3.55	131	0.0549	4.34
$(H_2PO_4^-)$	230	0.0565	7.20	235	0.0575	7.[illegible]
	230	0.0555	7.46	235	0.0572	7.18
	238	0.0552	7.81	235	0.0572	7.18
$(HPO_4^{=})$	189	0.0555	6.13	175	0.0562	5.53
	175	0.0565	5.48	175	0.0565	5.48
	175	0.0567	5.43	172	0.0575	5.30
$(PO_4^{\equiv})$	160	0.0712	3.16	132	0.0557	4.24
	147	0.0679	3.18	135	0.0562	4.27
	142	0.0628	3.60	130	0.0570	3.99

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TABLE E-2.

EQUILIBRIUM CONCENTRATIONS

Membrane Type: AR110DYG-067

Solution		Initial meq/1	Final meq/1	Δ meq/1
1	Cl^-	8.02	9.38 ; 9.38	+ 1.36 ; 1.36
	NO_3^-	3.22	1.90 ; 1.90	- 1.32 ; 1.32
2	Cl^-	6.70	9.20 ; 9.00	+ 2.50 ; 2.30
	$SO_4^=$	3.33	0.354 ; 0.375	- 2.98 ; 2.95
3	Cl^-	11.08	12.1 ; 12.1	+ 1.02 ; 1.02
	$H_2PO_4^-$	1.02	0.01 ; 0.01	- 1.01 ; 1.01
4	Cl^-	11.08	12.26 ; 12.26	+ 1.18 ; 1.18
	$HPO_4^=$	2.04	0.864 ; 0.862	- 1.18 ; 1.18
5	Cl^-	11.08	15.88 ; 16.3	+ 4.80 ; 5.20
	$PO_4^{\equiv}$	3.06	1.137 ; 1.107	- 1.92 ; 1.95

TABLE E-2 (continued)

EQUILIBRIUM CONCENTRATIONS

Membrane Type: AR110CZG-067

Solution		Initial meq/l	Final meq/l	Δ meq/l
1	Cl^-	8.02	9.45 ; 9.45	+ 1.43 ; 1.43
	NO_3^-	3.22	1.77 ; 1.77	- 1.45 ; 1.45
2	Cl^-	6.70	9.45 ; 9.65	+ 2.75 ; 2.95
	$SO_4^=$	3.33	0.521 ; 0.583	- 2.81 ; 2.75
3	Cl^-	11.08	12.1 ; 12.1	+ 1.02 ; 1.02
	$H_2PO_4^-$	1.02	0.02 ; 0.01	- 1.00 ; 1.01
4	Cl^-	11.08	12.18 ; 12.22	+ 1.10 ; 1.14
	$HPO_4^=$	2.04	0.948 ; 0.886	-1.094 ; 1.156
5	Cl^-	11.08	12.4	+ 1.3
	$PO_4^\equiv$	3.06	1.20	- 1.863

TABLE E-2 (continued)

EQUILIBRIUM CONCENTRATIONS

Membrane Type: AR111BZL-183

Solution		Initial meq/l	Final meq/l	Δ meq/l
1	Cl^-	8.02	10.52 ; 10.4	+ 2.50 ; 2.38
	NO_3^-	3.22	0.72 ; 0.82	- 2.50 ; 2.40
2	Cl^-	6.70	9.95 ; 9.95	+ 3.25 ; 3.25
	$SO_4^{=}$	3.33	0.542 ; 0.50	- 2.79 ; 2.83
3	Cl^-	11.08	12.10 ; 12.30	+ 1.02 ; 1.22
	$H_2PO_4^-$	1.02	0.102 ; 0.11	- 0.92 ; 0.91
4	Cl^-	11.08	11.42 ; 11.42	+ 0.34 ; 0.34
	$HPO_4^{=}$	2.04	1.708 ; 2.0	-0.334 ; 0.04
5	Cl^-	11.08	16.9 ; 17.4	+ 5.80 ; 6.30
	$PO_4^{\equiv}$	3.06	2.748 ; 2.88	-0.183 ; 0.315

TABLE E-2 (continued)

EQUILIBRIUM CONCENTRATIONS

Membrane Type: AR111GWL-183

Solution		Initial meq/l	Final meq/l	Δ meq/l
1	Cl^-	8.02	10.12 ; 10.1	+ 2.10 ; 2.08
	NO_3^-	3.22	1.12 ; 1.12	- 2.10 ; 2.10
2	Cl^-	6.70	9.5 ; 9.1	+ 2.80 ; 2.40
	$SO_4^{=}$	3.33	0.875 ; 1.122	-2.455 ; 2.21
3	Cl^-	11.08	11.95 ; 12.05	+ 0.87 ; 0.97
	$H_2PO_4^-$	1.02	0.00 ; 0.00	- 1.02 ; 1.02
4	Cl^-	11.08	11.54 ; 11.64	+ 0.45 ; 0.56
	$HPO_4^{=}$	2.04	1.58 ; 1.476	-0.462 ; 0.566
5	Cl^-	11.08	16.2 ; 15.9	+ 5.1 ; 4.8
	$PO_4^{\equiv}$	3.06	2.12 ; 2.14	-0.945 ; 0.921

TABLE E-3.

MEMBRANE SAMPLE CAPACITIES

	Membrane Type			
	AR110DYG-067	AR110CZG-067	AR111BZL-183	AR111GWL-183
Batch No.	347XA	236	184-A	Special
Capacity, meq/dry gram	2.58	1.8	1.8	1.32
Dry Cl^-/NO_3^- Sample wt	0.483	0.495	0.413	0.301
Sample meq	1.245	0.830	0.744	0.398
Dry $Cl^-/SO_4^=$ Sample wt	0.470	0.501	0.416	0.329
Sample meq	1.211	0.901	0.748	0.434
Dry $Cl^-/H_2PO_4^-$ Sample wt	0.454	0.493	0.410	0.325
Sample meq	1.170	0.886	0.737	0.429
Dry $Cl^-/HPO_4^=$ Sample wt	0.468	0.466	0.410	0.330
Sample meq	1.208	0.839	0.737	0.436
Dry $Cl^-/PO_4^≡$ Sample wt	0.410	0.415	0.405	0.326
Sample meq	1.060	0.746	0.728	0.430

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